CHE 4224

Spring 2018

2018 AICHE Design Report

“Dimethyl Ether for Transportation”

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Abstract

Our team designed a process to produce 250,000 gallons per day of dimethyl ether (DME) from a methanol feedstock for sale as a diesel fuel alternative. The proposed design utilizes four parallel reactors containing a gamma-alumina catalyst operating at 500 psia and 536°F to carry out methanol dehydration (1). The mixed reactor outlet is separated using two distillation operations to produce fuel-grade DME, methanol for recycle, and a wastewater stream.

$$2\ MeOH \leftrightarrow DME + H_2O \tag{1}$$

Table 1: Executive Summary Table

Executive Summary	
NPV	**$2.5 million**
ROR	**8.52%**
Initial Investment	**$70 million**
Investment Return Date	**2030**
Construction Timeline	**2018-2022**
Sale Price of DME/gallon	**$2.26**

The initial capital investment for the project is $70 million. Economic viability is highly contingent on the sale price of DME. As DME possesses half the energy density of diesel, DME must be sold at half the price of diesel to be competitive on the open market. With diesel around $3.00/gallon, DME should be targeted for $1.50/gallon, but this is not economically viable. At a sale price of $2.26 per gallon, the net present value (NPV) for the plant is $2.5 million and the rate of return (DCFROR) is 8.52%. If the DME sale price is reduced by 5%, however, the NPV drops to -$41 million.

Though there are a number of opportunities for increased revenue due to government initiatives to drive the adoption of renewable fuels, they are not expected to overcome the challenges of this market. Unless the current incentives are expanded, it is not recommended that this project is pursued into the detailed design phase.

Introduction

Diesel fuel is a power source harnessed by millions every day, as evidenced by the consistent rise in diesel consumption over the past decades (1). Despite its utility, diesel exhibits a number of limitations. Diesel fuel is not a clean-burning fuel, and thus leads to a significant release of incomplete combustion products. In addition to water vapor, carbon dioxide and carbon monoxide, soot particles released from vehicle tailpipes consist of polycyclic aromatic hydrocarbons and alkyl derivatives that are major contributors to air pollution (2). As such, much research has been put into finding cleaner and renewable alternatives for diesel fuel, including biodiesel, synthetic fuels, alcohols, hydrogen, and dimethyl ether (DME) (3).

Some have stated that DME is the most promising fuel of the future (4). It is a multipurpose, clean-burning fuel produced from coal, natural gas, or biomass (5). In recent testing, Volvo ranked DME ahead of other fuel alternatives, including CNG, LNG, ethanol, and biodiesel, in areas such as energy efficiency, cost, infrastructure, and environmental considerations (4). Furthermore, DME has potential to be a raw material for the production of hydrocarbons and chemicals, as a substitute propellant for chlorofluorocarbons (CFCs) as an aerosol (5), and as a solvent due to its low toxicity and high solubility of polar and non-polar compounds (6).

DME can be produced directly from synthesis gas, also known as syngas - a mixture of carbon monoxide and hydrogen - in a single process step using a bi-functional catalyst, or indirectly through a two-step reaction series with methanol as the intermediate product (7) (6). Since syngas can be manufactured from coal or biomass, DME can be produced through use of renewable resources, which through mass adoption can reduce the world's dependence on fossil fuels (7).

With the energy sector trending towards cleaner fuels and use of renewable resources, DME has the potential to be a lucrative revenue source in the near future. Our design team was tasked with putting together a preliminary design for the construction of a DME production plant in Lake Charles, Louisiana to take advantage of the projected market demand for this alternative fuel.

Market Analysis

The DME fuel market is a largely untapped potential source of revenue. Some companies, including Volvo as early as 2013, stated that it planned to pursue DME as a diesel alternative for its North America trucking industry (8). This occurred when the price of diesel was high, and while the enthusiasm cooled a bit as diesel prices dropped, DME is still seen as an upcoming market in the United States. Extensive DME field-testing in Sweden has shown potential for regional application, and California has been a leader in encouraging development of DME as a fuel in North America (9).

Barriers and Incentives to DME in Transportation

Though there are numerous environmental benefits to utilizing DME as a diesel fuel alternative, there are a considerable number of challenges to overcome before it enters the mainstream market. To convert a diesel engine to run on DME, the cylinder heads and fuel injection system must be replaced to handle the increase in fuel flow and decreased lubricity of the DME. Along with these engine upgrades, DME requires lubricant additives to protect the engine, which will increase the cost of production (10) (4) (11).

In addition, the fuel tank capacity must be increased in order to contain the 75 psig storage pressure of DME (9). Due to the differences in density and heating value, DME fuel provides a travel distance that is half as far as that of diesel fuel. Because of these differences DME tanks must be twice as large as diesel tanks for equivalent range (9). Furthermore, the fuel efficiency of a DME vehicle may be inhibited by the additional weight that the equivalent energy requires; one gallon of DME weighs 5.5 pounds to diesel's 7.5, but accounting for the two gallon equivalent leads to 11 pounds of DME per 7.5 pounds of diesel (9). To be competitive in the market, DME must be sold at approximately half the price of diesel, which does not take into account the up-front costs associated with converting diesel engines to DME.

Despite these challenges, projections on the DME market into the next decade suggest that the continuous increases in diesel consumption can benefit DME as a viable transport fuel. With this possibility, market growth is expected to be driven primarily by the transportation sector (12). A number of government regulation agencies are beginning to emphasize the need for clean fuel alternatives, which could lead to economic incentives for alternative fuels and more stringent regulation on conventional fuels. Several DME plants have been announced in lieu of this market outlook to cut the gap between supply and demand (12).

Oberon fuels, which has been a market leader for a number of years, states that their DME synthesis process has the benefit of generating renewable identification numbers (RINs) when the DME is produced from biogas (13). RINs act as a form of currency under the EPA's Renewable Fuel Standard Program and can be an alternative source of income for new DME synthesis processes. With this credit system in place, it is possible to earn profit in an otherwise challenging economic scenario by selling RINs to non-renewable fuel producers (14).

Additional government incentives include the SmartWay Transportation Partnership, the State Energy Program Funding, the Congestion Mitigation and Air Quality Improvement Program, and others that seek to encourage use of renewable and clean fuel sources (15). These programs provide increased means of income to a project and can improve the profitability of selected renewable fuels projects. Louisiana provides a 30% Fueling Infrastructure Tax Credit, which will cover 30% of the cost to convert a vehicle to alternative fuel (15). This will encourage adoption of DME and reduce the economic barrier within the trucking sector.

Historical Market Prices

Figure 1 displays historical data for diesel prices in the US Gulf Coast Region and Figure 2 compares the cost index value for methanol and diesel. The price spiked in 2008 and then dropped significantly in 2009, then grew to a plateau for 2012-2015 before experiencing another drop. The price is currently climbing and is about to reach $3/gallon. Since 1994, the price of diesel has increased an average of 7.2 cents per year, but has exhibited extreme variations throughout.

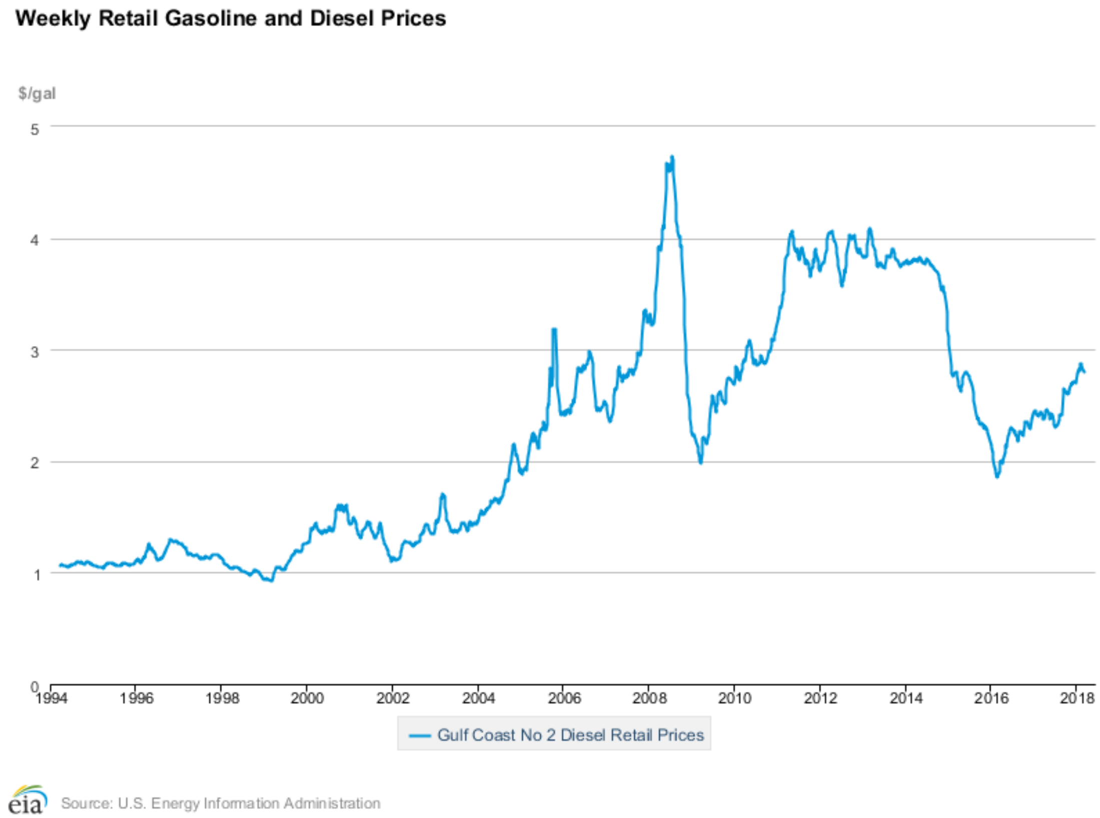


Figure 1: Diesel Fuel Cost per Gallon in the Gulf Coast Region *(16)*

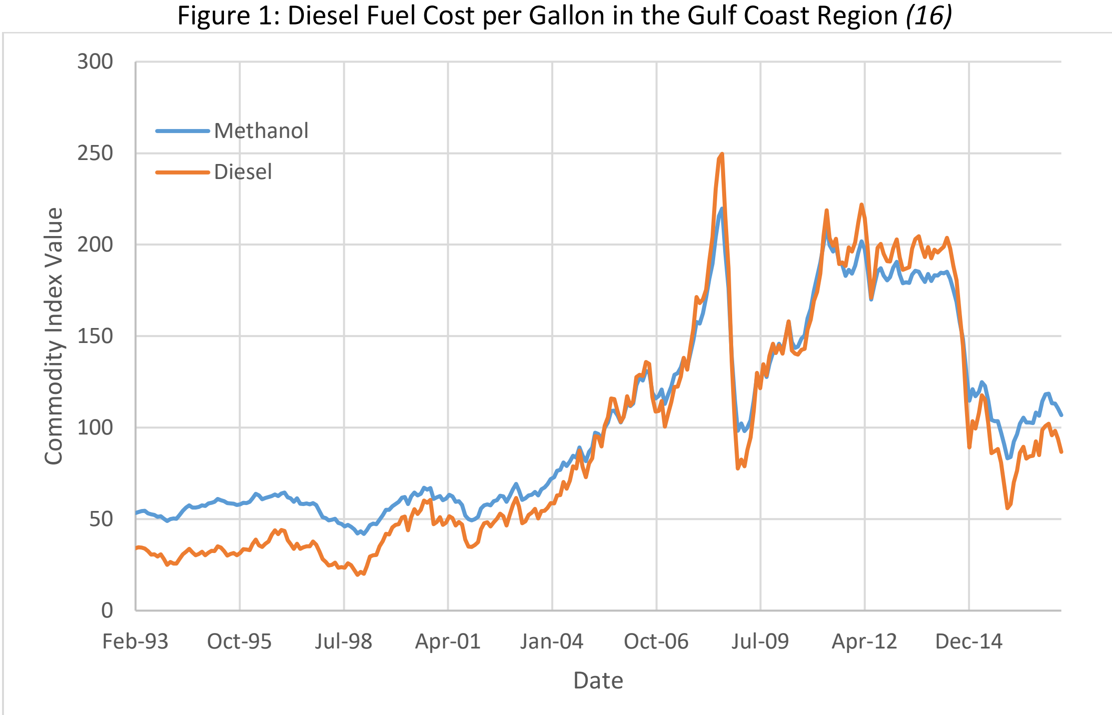


Figure 2: Methanol Cost per Gallon in the United States *(17)*

As seen in the previous figures, the price of methanol follows similar trends as diesel fuel. In order to be competitive on the open market, the sale price of DME must be comparable to half the price of diesel. With the national average price of diesel fuel at $2.99 per gallon and that of the Gulf Coast region at $2.79 per gallon (16), the DME must be sold for less than $1.50 per gallon. At this point, however, the price of methanol is $1.49 gallon (18), and as 1.1 gallons of methanol is required to produce 1 gallon of DME. At these prices, the process will not produce a profit.

With these market conditions, DME must be sold at a higher cost per energy equivalent of diesel. Additional measures must be included to encourage adoption of DME if market forces do not promote the transition currently. Further suggestions for improving the profitability of the process can be found on page 90.

Project Charter

The proposed plant design must be able to produce 250,000 gallons per day of DME utilizing the indirect synthesis method with methanol as the feed material. This production rate should be able to support a shipping region of 2,000 trucks per day. Methanol is to be delivered by railcar from the Lake Charles MeOH plant. The proposed design must be able to turn down to 50% production capacity while maintaining efficient plant operation. The DME product must meet ISO DME Fuel Plant Gate Standards (Table 1) and must also include a 900 ppm concentration of lubricant additives (Table 2). All designs must also be environmentally conscious and include inherent, passive, and active safety measures.

Economic measures will be determined using a 20-year plant life and a discount rate of 8%. Capital costs and operational costs must be accounted for in the economic analysis, but the costs for the DME truck filling station can be excluded. The production costs of DME and the required incentives to motivate transition from diesel to DME should be included in analysis, and an overall risk assessment for technical, business, and safety aspects should be undertaken. Recommendations must include ways to reduce the carbon footprint of the process, options to monetize waste products, and further safety and environmental impact improvements.

Table 1: Requirements and Design Specs for DME

Property	Requirement	Design Specs
DME mass fraction	0.985 minimum	0.9995
Methanol mass fraction	0.0005 maximum	0.0005
Water mass fraction	0.0003 maximum	0
Methyl Formate mass fraction	report	0

Table 2: Lubricant Specs

Lubricant spec	Value	Units
Required concentration	**900**	**ppm**
Bulk cost	**$1.65**	**$/lb**
Shipping cost	**$0.02**	**$/gallon**

The plant may utilize any process, but should follow the reaction of Methanol to DME shown below (1).

$$2\mathrm{MeOH} \leftrightarrow \mathrm{DME} + \mathrm{H_2O} \tag{1}$$

The reaction is typically catalyzed by a heterogeneous acid catalyst, and it exhibits exothermic and equilibrium characteristics. One method for production is use of a high-temperature, gas-phase process and a gamma alumina oxide (γ-Al_2O_3) catalyst, while another option includes a liquid-phase reaction over a super-acid polymer resin. While the rate of reaction increases with an increase in temperature, DME conversion and selectivity is favored at lower temperatures. Kinetic literature values and equations were provided for both the high temperature and the low temperature processes. In addition, thermodynamic and equilibrium data for methanol, DME, and water mixtures was provided. The rate of reaction equation is displayed (below) and Table 3 contains kinetic data for the reaction.

$$Rate\left(\frac{kgmole}{m3 * s}\right) = \frac{k \times [MeOH]^m - k' \times [H2O][DME]}{(1 + K1 \times [Methanol]^{(f1\_Methanol\_exp)} \times [H2O]^{(f1\_H2O\_exp)} + K2 \times [H2O]^{(f2\_H2O\_exp)})}$$

$$\mathrm{k, k'} = \mathrm{A} \times \mathrm{e}^{-\frac{\mathrm{E}}{\mathrm{RT}}} \tag{2}$$

$$\mathrm{K1, K2} \ldots = \mathrm{A1} \times \mathrm{e}^{-\frac{\mathrm{E1}}{\mathrm{RT}}} \tag{3}$$

$[Methanol]$	Mole fraction of methanol
$[H2O]$	Mole fraction of water
$[DME]$	Mole fraction of DME

Each catalyst has distinct limitations to implementation. The low temperature catalyst will decompose and release sulfate groups, which form sulfuric acid in the presence of water. While this is negligible at temperatures lower than 120°C, it leads to a 15.9% weight loss at 160°C. Reactors are typically thirty feet tall or shorter, and it is desirable to maintain a pressure drop of less than 15 psig to prevent bead compression and screen blinding. The high temperature catalyst is effective up to 400°C, after which it will degrade.

The proposed plant is to be constructed close to the Lake Charles Methanol Plant and in close proximity to railways, pipelines, and highways. Since the property has not yet been acquired, recommendations for relocation are acceptable. The plant will include a filling station, and safety and environmental hazard analysis must account for the presence of trucks on the premises as well as the process itself.

Table 3: Kinetic data for both the high and low temperature processes

Catalyst Reaction Kinetic Parameters and Data	High Temperature Gas Phase (Gamma Alumina catalyst)	Low temperature Liquid Phase (high acid resin i.e. Amberlyst 35)
Heat Of Reaction, kJ/kgmole	-11,712	-11,712
A forward, kgmole/m3-s	1.0626x10^6	1.2457x10^11
E forward, kJ/kgmole	65,633	98,000
m Forward Reaction Order	2	0
A' reverse, kgmole/m3-s	1.4677x107	(no reverse rxn data)
E' reverse, kJ/kgmole	88,994	(Correlation limited from pure to 5mol/litre MeOH)
A1	0.5366	1.565x10-3
E1, kJ/kgmole	-3,450	-24,643
f1_Methanol _exp	0.5	-1
f1_H2O_exp	0	1
A2	4.50x10-2	n/a
E2, kJ/kgmole	-9,395	n/a
f2_H2O_exp	1	n/a
n, Denominator Exponent	4	2
Maximum Temp, °C	400 (catalyst deactivation), Ref: Turton	150 (catalyst limit) ref: DOW data sheet
Price	$4.65/lb - $5.25/lb.	$15 / lb in large quantities
Bulk Density (typical), gm/cm3	0.882	0.607
Material Density (typical), gm/cm3	1.47	1.504
Void Fraction	0.4	0.6
Life	9 to 12 months	unknown

Process Flow Diagram and Material Balances

Process Selection and Justification

The design team selected the high-temperature reaction with the ɣ-Al_2O_3 catalyst for the process. Table 4 displays some of the design considerations.

Table 4: Comparison of process catalyst

Parameter	ɣ-Al2O3	Amberlyst-35
Reaction phase	**Vapor**	**Liquid**
Maximum Temperature (°C)	**400**	**160**
Catalyst life (months)	**9 to 12**	**Varies with process conditions**
Water inhibition	**Significant**	**Minimal**
Reaction rate	**High**	**Low**
Purchase price	**$4.65/lb - $5.25/lb.**	**$15 / lb in large quantities**
Pressure drop considerations	**Pressure drop lowers reactant concentrations**	**Must be less than 30 psi**
Industrial adoption rates	**High**	**Low**
Heat of reaction	**Increases Reaction Rate**	**Must be accounted for to prevent catalyst inactivation**

While each catalyst has its own benefits, ultimately the ɣ-Al_2O_3 catalyst was chosen because of its proven track-record in industrial processes as evidenced by its high adoption rate, and that the heat of reaction assists the process by increasing the rate of reaction with minimal risk of catalyst degradation. Preliminary calculations into each reaction type yielded information suggesting that the cooling systems required to maintain reactor temperature, chilled water utilities, and reactor sizes needed to utilize the Amberlyst-35 catalyst were not as economically attractive as the proven alumina catalyst method. Other catalysts considered include a number of zeolite derivatives that each had distinct properties that provided advantages in selectivity and stability at high temperatures (19), but were also dismissed due to cost factors and availability of kinetic data.

Upon selection of the catalyst, the method to utilize the catalyst was determined. The majority of industrial synthesis processes utilize a packed bed reactor that produces a mixed outlet of methanol, water, and DME that is subsequently sent to a series of distillation columns for separation (7) (20). In the first column DME is removed from the mixture as distillate and sent to a mixing unit for lubrication before being sent to the truck filling station. The bottoms product of the first column then functions as the feed to the second to separate the remaining water from methanol. The distillate methanol stream is recycled back to the head of the process to be sent through the reactor again, minimizing waste and maximizing profit.

Another process that has garnered attention recently for the production of DME from methanol is catalytic distillation. In catalytic distillation, the reactants are fed into a distillation column containing both trays for separation and catalyst for reaction. As the methanol reacts on the catalyst, the newly formed DME moves preferentially up the column due to its higher volatility and exits the column in the distillate, while the less volatile water exits the column in the bottom product. While catalytic distillation provides cost benefits through combining the reaction and separation phases, this process was not chosen due to its unproven status in the field of DME production and challenges in operation and control.

Material Balance

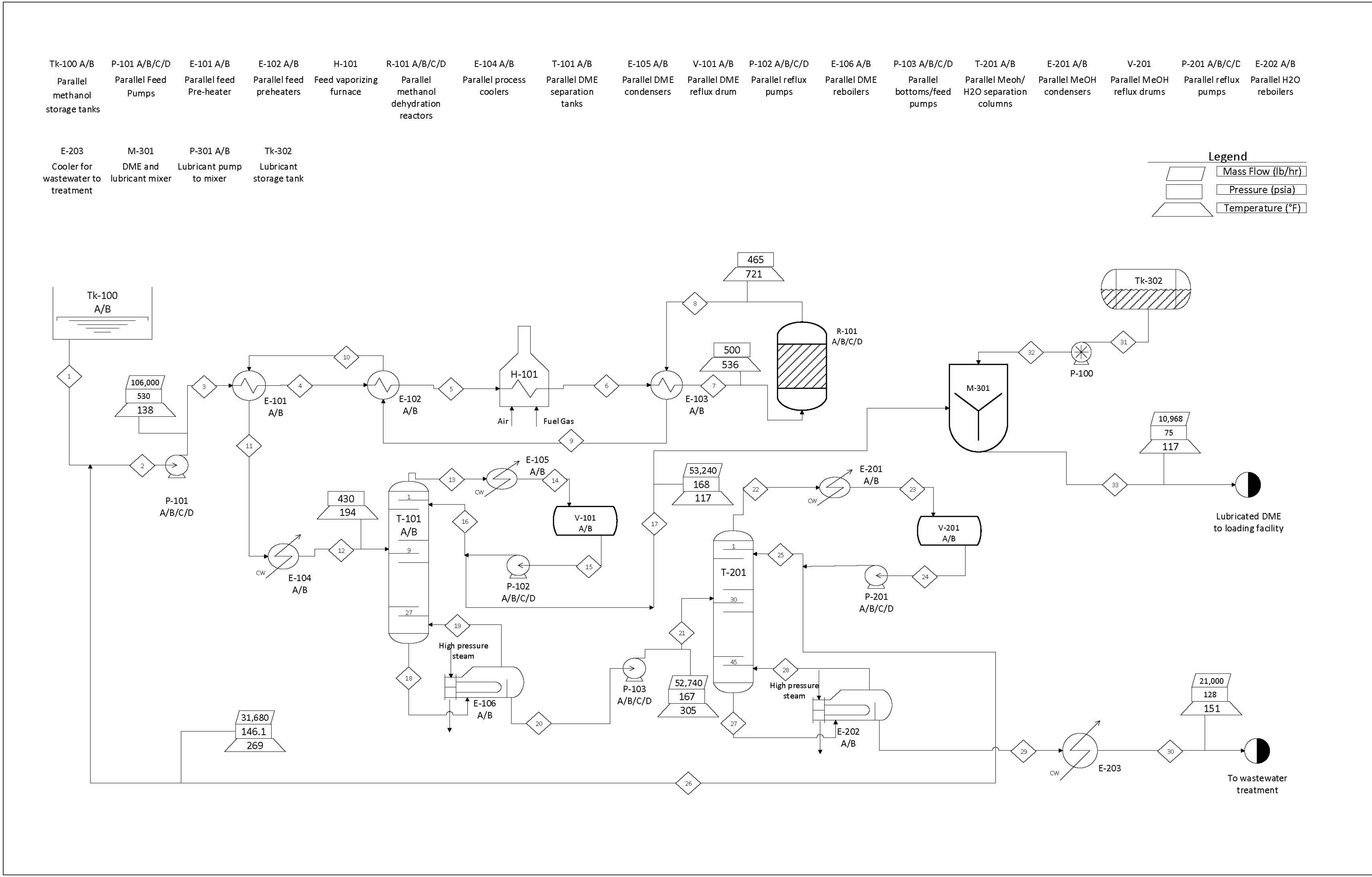
The overall mass balance on the system was determined from the Aspen HYSYS simulation. The inlet flow rate of methanol and the wastewater outlet rate were calculated from the required DME production rate. These values were specified in the feed stream and the distillation column product streams. As these values were specified, the overall mass balance is accurate to less than a tenth of a percent difference. Table 5 displays the results.

Table 5: Overall system mass balance

Stream	Methanol Feed	DME Outlet	Wastewater Outlet
Mass Flow (lbm/hr)	**74,296**	**53,248**	**21,060**
Mass Balance	**74,296 -(53,248 + 21,060)=-6**		
Percent Difference (%)	**0.016**		

Process Flow Diagram and Stream Table

The following pages display the process flow diagram (PFD) and Stream Table for the proposed process.



Tk-100 A/B
Parallel methanol storage tanks
P-101 A/B/C/D
Parallel Feed Pumps
E-101 A/B
Parallel feed Pre-heater
E-102 A/B
Parallel feed preheaters
H-101
Feed vaporizing furnace
R-101 A/B/C/D
Parallel methanol dehydration reactors
E-104 A/B
Parallel process coolers
T-101 A/B
Parallel DME separation tanks
E-105 A/B
Parallel DME condensers
V-101 A/B
Parallel DME reflux drum
P-102 A/B/C/D
Parallel reflux pumps
E-106 A/B
Parallel DME reboilers
P-103 A/B/C/D
Parallel bottoms/feed pumps
T-201 A/B
Parallel Meoh/ H2O separation columns
E-201 A/B
Parallel MeOH condensers
V-201
Parallel MeOH reflux drums
P-201 A/B/C/D
Parallel reflux pumps
E-202 A/B
Parallel H2O reboilers
E-203
Cooler for wastewater to treatment
M-301
DME and lubricant mixer
P-301 A/B
Lubricant pump to mixer
Tk-302
Lubricant storage tank
Legend
Mass Flow (lb/hr)
Pressure (psia)
Temperature (°F)
Tk-100 A/B
106,000
530
138
P-101 A/B/C/D
E-101 A/B
E-102 A/B
H-101
Air
Fuel Gas
E-103 A/B
465
721
500
536
R-101 A/B/C/D
Tk-302
P-100
M-301
10,968
75
117
Lubricated DME to loading facility
E-104 A/B
CW
430
194
T-101 A/B
E-105 A/B
V-101 A/B
P-102 A/B/C/D
High pressure steam
E-106 A/B
53,240
168
117
E-201 A/B
V-201 A/B
T-201
P-201 A/B/C/D
P-103 A/B/C/D
52,740
167
305
High pressure steam
E-202 A/B
21,000
128
151
E-203
To wastewater treatment
31,680
146.1
269
1
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Table 6: Stream Table

Stream I.D.	Description	Temperature (°F)	Pressure (psia)	Total Mass Flow (lb/hr)	Mass Frac. MeOH	Mass frac. DME	Mass frac. H2O
1	Methanol feed from storage tanks	77.0	14.7	74,300	0.9990	0.0000	0.0010
2	Combined feed including recycled methanol	137.6	14.7	106,000	0.9985	0.0004	0.0111
3	Pressurized feed	140.4	530.0	106,000	0.9885	0.0040	0.0111
4	Partially heated feed	293.4	524.0	106,000	0.9885	0.0040	0.0111
5	Feed entering furnace for phase change	379.4	516.0	106,000	0.9885	0.0040	0.0111
6	Vapor feed into final preheater	379.4	516.0	106,000	0.9885	0.0040	0.0111
7	Heated feed into reactor	536.0	500.0	106,000	0.9885	0.0040	0.0111
8	Reacted process stream into first cooling HEX	721.4	465.0	106,000	0.2902	0.5024	0.2074
9	Partially cooled process stream into 2nd cooling HEX	582.6	456.3	106,000	0.2902	0.5024	0.2074
10	Process stream entering 3rd cooling HEX	390.2	447.5	106,000	0.2902	0.5024	0.2074
11	Process stream into CW HEX for condensation	358.9	438.8	106,000	0.2902	0.5024	0.2074
12	Condensed process stream into first separation tower	194.4	430.0	106,000	0.2902	0.5024	0.2074
13	DME overhead vapor stream into condenser	117.6	153.7	85,400	0.0005	0.9995	0.0000
14	condensed DME stream into reflux drum	117.3	153.7	85,400	0.0005	0.9995	0.0000
15	Liquid DME stream into reflux pump	117.3	153.7	85,400	0.0005	0.9995	0.0000
16	DME stream into column as reflux	117.3	167.5	31,400	0.0005	0.9995	0.0000
17	DME stream into lubricant mixer	117.0	167.5	53,240	0.0005	0.9995	0.0000
18	Bottoms liquid stream into reboiler	300.5	156.0	86,900	0.6708	0.0011	0.3280
19	Boil-up stream entering column	305.2	156.0	34,200	0.8068	0.0022	0.1910
20	Bottoms stream entering 2nd column feed pump	305.2	156.0	52,700	0.5828	0.0004	0.4168
21	Feed stream pumped into 2nd column	305.2	166.7	52,700	0.5828	0.0004	0.4168
22	Methanol rich overhead vapor stream into condenser	270.7	124.7	63,900	0.9643	0.0007	0.0350
23	Condensed methanol stream into reflux drum	269.2	124.7	63,900	0.9643	0.0007	0.0350
24	Methanol stream into reflux pump	269.2	124.7	63,900	0.9643	0.0007	0.0350
25	Methanol rich stream entering column as reflux	269.2	146.1	32,300	0.9643	0.0007	0.0350
26	Methanol rich stream recycled into original methanol feed	269.2	146.1	31,700	0.9643	0.0007	0.0350
27	Bottoms water rich stream into reboiler	341.3	128.4	52,200	0.0297	0.0000	0.9703
28	Water rich stream entering column as boil up	344.8	128.4	31,100	0.0438	0.0000	0.9562
29	Bottoms product wastewater into cooler	344.8	128.4	21,100	0.0089	0.0000	0.9911
30	Wastewater stream cooled to treatment plant	150.8	128.0	21,100	0.0089	0.0000	0.9911
31	Lubricant stream entering mixer pump	77.0	14.7	48	0.0000	0.0000	0.0000
32	Pumped lubricant into mixer	77.0	75.0	48	0.0000	0.0000	0.0000
33	DME rich stream mixed with lubricant to loading facility	117.0	75.0	53,200	0.0005	0.9995	0.0000

Process Description

Block Flow Diagram

The overall process can be reduced to a reaction followed by a series of separations. As shown in Figure 3, methanol fed to the system enters a reactor, where it partially reacts to form DME and water. The mixed outlet feeds to a distillation column, which separates the DME from the water and methanol. The remaining methanol and water feeds into a second distillation column, which separates the water from the methanol. The methanol is recycled to the head of the process for further reacting, while the water is sent to the wastewater treatment plant to be discharged into the environment.

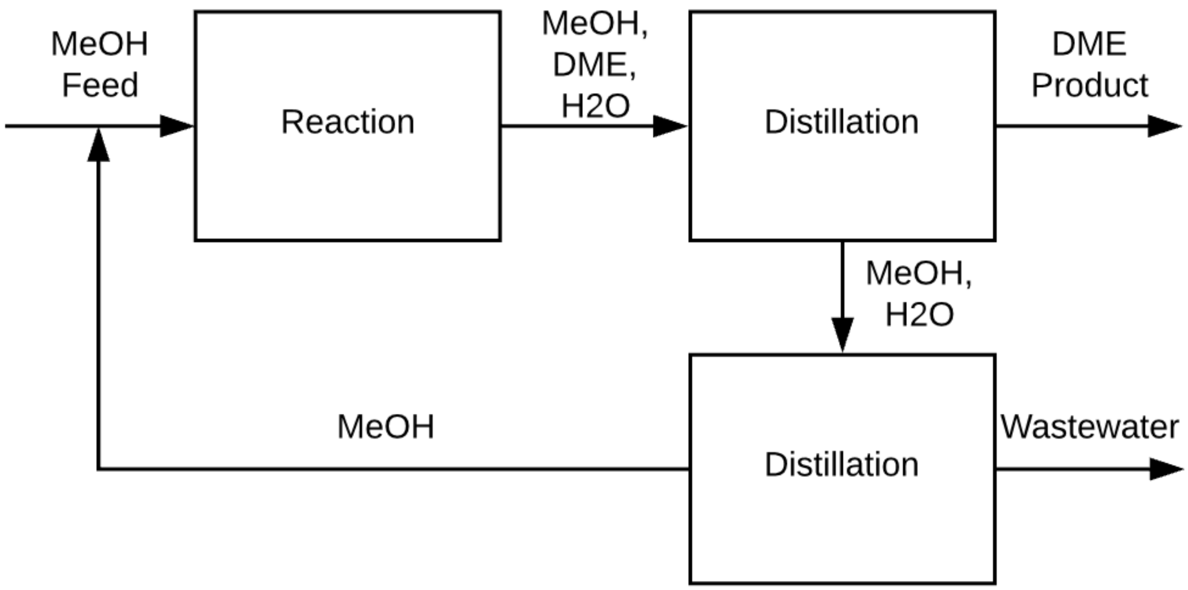


Figure 3: Block Flow Diagram for the proposed DME synthesis process

Detailed Process Description

Methanol Inlet

The methanol for the process is received from railcars and is stored in two stainless steel floating roof storage tanks, Tk-100 A/B. These were selected for their ability to maintain the purity of the methanol over long periods of time, as they mitigate the risk of water from the atmosphere entering. In addition, the stainless steel material is required for maintaining the structural integrity of the tanks, as methanol can corrode carbon steel over the course of years. A carbon steel tank would have a lifetime of 4-5 years, while the stainless steel tank will last for the lifetime of the project.

The recycle stream from the second distillation column (Stream 26) combines with the inlet methanol stream exiting the storage tanks (Stream 1) and raises the temperature to 137.5°F before entering P-101 A/B/C/D, the parallel, multistage centrifugal feed pumps. Each pump is designed to be able to operate alone and produce half of the required daily flow rate, or 53,000 lb_m/hr of the total 106,000 lb_m/hr. These pump types were selected because of their ability to produce the 530 psig pressure required for the process without the limitations of a positive displacement pump. In addition, as there are two heat exchangers in parallel operation for each section of the heat integration system, any of the pumps can operate alone to meet the specification for 50% turndown given in the project charter.

Stream Preheating

The methanol exits the pumps and enters a series of heat exchangers. Each heat exchanger (excluding E-104) is designed for only a single type of heat exchange, as this eliminates the challenges of operating a heat exchanger that includes both sensible and latent heat transfer in the same process stream. Furthermore, each heat exchanging operation was designed as two parallel units (A/B) to allow for 50% turndown and while maintaining design operation. This configuration was also less costly than if the transfer area required was sized as a single exchanger.

The first heat exchanger (E-101 A/B) that the methanol feed enters accounts for sensible heat transfer on the tube side to preheat the reactant from 137°F to 293°F, and then subsequently into a second heat exchanger (E-102 A/B) to raise the temperature from 293°F to 379°F. Each of these heat exchangers was designed as an AEL TEMA type because of the economic value of the fixed tube design and the process flexibility of the E type shell, which is good for both latent and sensible heat transfer. These heat exchangers were designed for a pressure of 580 psig, which includes a safety factor of over 50 psi. They were separated into two distinct heat exchangers to separate the sensible heat transfer and phase change in the shell side stream. Further information is provided on page 28.

The methanol exits the E-102 A/B as a saturated liquid at 379°F and combines with its parallel stream to enter the preheating furnace, H-101. Whereas the heat exchangers were sized so that two could operate in parallel, a single furnace was far more cost-efficient than parallel operation, and thus the total flow is heated in the single unit. Within the furnace, the methanol stream is completely vaporized and exits the furnace as a saturated vapor at 380°F. While the preheating furnace is highly effective at transferring the heat required to the stream, it does incur a significant cost to operate due to the fuel gas requirement. Despite this drawback, heat integration challenges (see page 28) eliminated the possibility of total heat integration, and the process steam was not hot enough to provide the energy required. As such, the furnace was selected because of its utility in spite of the cost.

Once Stream 6 exits H-101 as a saturated vapor, it splits into two parallel streams again and enters E-103 A/B, which increases the feed temperature to the 536°F inlet temperature required (Stream 7). This exchanger is also an AEL design, taking advantage of the cost and versatility of the design. As before, the reactant stream flows through the tube side of the unit while the shell side is reserved for cooling the reactor product. At the outlet of E-103 A/B, the pressure of the stream has reached 500 psia, the design pressure for the reactor. Page 28 displays the details of the HEX operation.

Reactors

Stream 7 splits into four equal flow streams and enters R-101 A/B/C/D, four parallel packed-bed reactors. Vapor enters the vertical reactors at 500 psia and 536°F and exits the bottom at 465 psia and 720°F. Within the reactor, methanol dehydration (1) occurs on the surface of the catalyst. R-101 is designed as a vertical pressure vessel with 80 3” nominal tubes inside of which houses the gamma-alumina catalyst. The mass of the catalyst in each reactor is 3,520 pounds. At these process conditions, methanol conversion is 70%, which was determined due to its balance of operability and the minimization of overall system size and expense. Page 24 provides more details on the reactor sizing calculations.

Product Stream Cooling

Stream 8 exits R-101 A/B/C/D at 720°F and 465 psia and a molar composition of 36.6% water, 34.6% DME, and 28.8% water. This mixed product splits into two streams again and enters the shell side of E-103 A/B and cools to 583°F. Immediately upon exiting this heat operation, Stream 9 enters the shell side of E-102 A/B where it further cools until it becomes a saturated vapor at 390°F. Once it becomes a saturated vapor, it enters E-101 A/B (Stream 10) where the fluid begins to condense. Since the stream is not a pure component, sensible heat transfer occurs along with latent heat transfer, and Stream 11 exits E-101 A/B as a two-phase mixture at 359°F. Additional information can be found on Page 28.

This two-phase mixture immediately enters the shell side of a final pair of parallel AEL heat exchangers. Cooling water flows through the tube side to absorb energy and drop the temperature of the product stream to 194°F. The cooling water enters at 90°F and exits at 120°F, which requires a flow rate of 1,040 gpm. Cooling water temperatures were estimated through environmental wet bulb temperatures that could be experienced in the region and a desire to mitigate bacterial growth within the cooling towers and process equipment. In a similar fashion as before, the AEL TEMA type was selected for its low cost and versatility. Once the product stream exits the shell side of the exchanger, its pressure is 430 psia and is subsequently sent to the separation portion of the process.

DME Isolation

The mixed product stream enters the first of two distillation units. Distillation was selected as the preferred method of separation because it is proven technology and has been successfully implemented in similar processes numerous times before. The relative volatility of the components in the mixture is high, meaning that a smaller number of trays and lower reflux ratio is required than would be in a mixture with lower relative volatilities. Though there are other separation means for these chemicals, none approach distillation in range of application or length of field success.

Each separation is designed with two towers operating in parallel, which is justified using the same operability at 50% turndown reasoning and lower capital cost arguments as applied to the heat exchangers. Each of the DME isolation towers has the feed enter at tray 9 out of 27, numbered from the top. Tray spacing was set at 2 feet for ease of maintenance. The column is specified to produce a distillate flow rate of 26,600 lb_m/hr of 99.95 wt.% DME, both of which are requirements stated in the project charter. While this is above the 99.85% specification required by the ASTM standard, higher DME purity is required to meet the water composition requirement, which is a maximum of 0.0005 mass fraction water. A mixture of water and methanol exits the reboiler through the bottom product pump.

The column operates at a reflux ratio of 0.5723, and the condenser pressure is set at 153.7 psia. The operating pressure was set by the minimum approach temperature for the cooling water on the condenser, with heuristics requiring a 30° condenser approach temperature between the cooling water inlet and the saturation temperature of the overhead product (21). The reboiler pressure was estimated at 156 psia by adding 0.1 psia pressure drop for each tray in the column to the condenser pressure. The reboiler operates at 305°F. This combination of reflux ratio, number of trays, and feed tray was found to be the lowest 20-year PWC for the combination of condenser, tower, and reboiler. Sizing calculations are shown on Page 33, while costing estimates are shown on Page 65.

The distillate stream containing the DME is sent to an inline mixing unit where lubricant is added to produce a final concentration of 900 ppm. The lubricant is stored in a tank containing a 30-day supply of lubricant to ensure that production capacity can be met for the same length of time that the methanol feed tanks can supply the process. The lubricant is fed into the baffled mixing vessel using a metering pump to precisely control the rate at which the lubricant is added and ensure product quality. An inline mixer was selected because of its minimal operating costs, and it provides continuous homogenization and efficient mass transfer. The DME exiting this mixer travels to the truck filling station to be sold for profit.

Methanol Recovery

A water/methanol mixture exits the reboiler of the first column at 156 psia, 305°F, and a flow rate of 26,400 lb_m/hr. This stream enters a pump that lifts the stream up to the 30th of 45 trays numbered from the top of the second separation unit. As with the previous column, the condenser pressure, determined to be 124.7psia, is set by the temperature of available cooling water, and the reboiler pressure was calculated by adding 0.1 psia pressure drop per tray. At these pressures the condenser temperature was 269°F while the reboiler was 344.8°F.

The column was specified using the wastewater flow rate specified by the overall system material balance and a methanol composition specified by the lower flammability limit (LFL) of the vapor above the water. The LFL for methanol is 6.7% by volume, so the molar composition of the wastewater was specified to be 0.995, which correlates to a vapor phase volume fraction of 5.4%. At these conditions, the water flow rate is 10,500 lb_m/hr. With these specifications, the lowest PWC column contained 45 trays in a 3-feet diameter column with 1.75 feet tray spacing and operated at a reflux ratio of 0.9942. The methanol recovery towers were sized and costed using the calculations described on page 65.

The distillate contains 93.4% methanol and is returned to the head of the process in a recycle stream at 125 psia and 269°F, where it is immediately pumped back through the system. This recycle design ensures maximum recovery of energy from the process stream and reduces the energy that needs to be added to preheat the reactant stream.

The wastewater from the bottom of the two parallel columns is discharged to a gravity sewer that is to be tied in to the existing city sewers in the area. The water flow rate is approximately 48 gallons per minute, but assuming the city has a 10-inch sewer with a 0.3% slope located in the proximity to the plant, the 48 gpm is only 21% of the total flow capacity (22), suggesting that the sewer could handle the increased wastewater capacity without modification to the collection system.

Energy Balance and Utility Requirements

Table 7: Energy Balance and Utility Requirement Table

Equipment	Q (Btu/h)	Quantity minus sparing	Overall Q	Purpose	Total utility requirment
P-101	2.27E+05	2	4.53E+05	Takes 2.27E+05 Btu/hr of electricty from the driver to increase the pressure of stream 2 from 14.7 psia to 500 psia	132.8 kW
E-101	7.70E+06	2	1.54E+07	Takes 7.70E+06 Btu/hr of energy to heat stream 3 from 137.5 F to 293.4 F. This heat was produced by cooling and condensing stream 10 from 390.2 F to 358.9 F	N/A
E-102	5.14E+06	2	1.03E+07	Takes 5.14E+06 Btu/hr of heat to heat stream 4 from 293.4 F to 379.4 F. This heat was produced by cooling stream 9 from 582.6 F to 390.2F	N/A
H-101	3.47E+07	1	3.47E+07	Takes 3.47E+07 Btu/hr of energy to vaporize stream 5 using fuel gas	3.40E+08 ft3/year of Natural gas
E-103	3.72E+06	2	7.43E+06	Takes 3.72E+06 Btu/hr of energy to heat stream 6 from 379.6 F to 536 F. This heat was provided by cooling stream 8 from 721.4 F to 582.6 F.	N/A
E-104	1.77E+07	2	3.54E+07	Takes 1.77E+07 Btu/hr of energy to cool Stream 11 from 358.9 F down to 199.4 F by heating cooling water from 89.6 F up to 120.2 F, Cooling water is supplied at 8.65 E+03 gpm.	1.73 E+04 gpm of cooling water
E-106	8.31E+06	2	1.66E+07	Takes 8.31E+06 Btu/hr of energy to vaportize stream 18. The energy is supplied by low pressure steam at a rate of 4.02 lb/sec.	8.04 lb/sec of low pressure steam
E-105	7.15E+06	2	1.43E+07	Takes 7.15E+06 Btu/hr of energy to condense stream 13. This is achieved by heating cooling water from 84.2 F to 98.6 F, cooling water is supplied at 7.42 E+03 gpm	1.48 E+04 gpm of cooling water
P-102	7.57E+03	2	1.51E+04	Takes 7.57E+03 Btu/hr of electricity from the driver to increase the pressure of stream 15 from 153.7 psia to 169.5 psia in order to pump reflux back into tower and to pump DME product into M-301	4.44 kW
P-103	2.79E+03	2	5.58E+03	Takes 2.79E+03 Btu/hr of electricty from the driver to increase the pressure of stream 20 from 156.4 psia to 164.8 psia in order to pump the bottoms of T-101 into T-201	1.64 kW
E-202	1.32E+07	2	2.64E+07	Takes 1.32E+07 Btu/hr of energy to vaporize stream 27. This heat was produced by supplying low pressure steam at a rate of 3.99 lb/sec	7.98 lb/sec of low pressure steam
E-201	1.34E+07	2	2.67E+07	Takes 1.34E+07 Btu/hr of energy to condense strean 22. This is achieved by heating cooling water from 84.2 F to 98.6 F, cooling water is supplied at 1.38E+04 gpm.	2.76E+04 gpm of cooling water
P-201	7.92E+03	2	1.58E+04	Takes 7.92E+03 Btu/hr of electricty from the driver to increase the pressure of stream 24 from 124.7 psia to 150.3 psia to pump the reflux into the column and to pump recycle back into system	4.64 kW
P-301	3.07E+01	1	3.07E+01	Takes 3.07E+01 Btu/hr of electricty from the driver to increase the pressure of stream 31 from 14.7 psia to 75 psia in order to pump lubricant into the mixing tank	.018 kW
E-203	1.98E+06	2	3.96E+06	Takes 1.98 E+06 Btu/hr to cool stream 30 from 173.8 F to 66 F, by heating cooling water from 32 F to 49 F at a rate of 962 gpm	1.93 E+03 gpm of cooling water

Equipment List and Unit Descriptions

R-101 A/B/C/D

The total process flow was split between four parallel packed bed reactors. Four reactors were utilized for their benefit to operability and turndown. Operating two reactors in parallel provides flexibility to maintain design point reactor operation even in turndown scenarios, and having four in parallel rather than two allows for smaller, less expensive vessels and the flexibility to operate at 75% capacity while the catalyst in one reactor is replaced. The reaction kinetics for the process were modeled in Polymath. The kinetic parameters provided in the project charter were loaded into the differential equation (DE) solver in terms of conversion and volume of catalyst. Side reactions were considered negligible at the process conditions.

A	Methanol
B	DME
C	Water

Table 8: Variables for Catalytic Reaction Analysis

Designation	Variable
C_i	**Concentration of material i**
X	**Conversion of methanol**
F_{ao}	**Flow rate of methanol at inlet**
r_a	**rate of reaction with respect to material a**
R	**Gas constant**
W	**Catalyst weight**
C_{ao}	**Concentration of methanol at inlet**
T	**Temperature**
T_o	**Temperature at inlet**
y	**Pressure ratio (between 0 and 1)**
α	**Geometric pressure drop coefficient**
θ	**ratio of inlet feeds to methanol**
C_{pi}	**Heat capacity of i**
ΔC_p	**Change in constant pressure heat capacity**
x_{io}	**Inlet mole fraction of i**

Design Equation	$$\frac{dX}{dW} = -\frac{r_a}{F_{ao}}$$ $$X(0) = 0$$	4
Rate Equation	$$-r_a = \frac{kC_a{}^{m} - k'C_bC_c}{(1 + K1C_a{}^{(f1\_Methanol\_exp)}C_c{}^{(f1\_H2O\_exp)} + K2C_c{}^{(f2\_H2O\_exp)})}$$	5
Rate Constants	$$k, k' = A \times e^{-\frac{E}{RT}}$$	6
Equilibrium Constants	$$K1, K2 \ldots = A1 \times e^{-\frac{E1}{RT}}$$	7
Stoichiometry	$$C_a = C_{ao}(1 - X)\left(\frac{T_o}{T}\right)y$$	8
	$$C_b, C_c = \frac{1}{2}C_{ao}(X)\left(\frac{T_o}{T}\right)y$$	9
Pressure Dependence	$$\frac{dy}{dW} = \frac{-\alpha}{2y}\left(\frac{T_o}{T}\right)$$ $$y(0) = 1$$	10
Temperature Dependence	$$T = \frac{X(-\Delta H_{rxn}) + T_o \sum \theta_i C_{p,i} + 298\Delta C_p X}{\sum \theta_i C_{p,i} + X\Delta C_p}$$	11
	$$\Delta C_p = .5(C_{pb} + C_{pc}) - C_{pa}$$	12
	$$\sum \theta_i C_{p,i} = C_{pa} + \frac{x_{bo}}{x_{ao}} C_{pb}$$	13

Constants and kinetic data was acquired through the project charter and additional scholarly resources. The pressure drop coefficient is based on the size of the particles and the geometry of the system; the value used is based on a derivation from Ahmed et. al. (20). The vapor phase heat capacities for each of the chemicals was estimated at approximately 650 K to eliminate self-referencing in the temperature dependence equation. All constants used in analysis are displayed in Table 9.

Table 9: Constants and Values for Catalytic Reaction Analysis

Parameter	Value
Heat Of Reaction, kJ/kgmole	**-11,712**
A forward, kgmole/m^3-s	**1.0626x10^6**
E forward, kJ/kgmole	**65,633**
m Forward Reaction Order	**2**
A’ reverse, kgmole/m^3-s	**1.4677x10^7**
E’ reverse, kJ/kgmole	**88,994**
A1	**0.5366**
E1, kJ/kgmole	**-3,450**
f1_Methanol _exp	**0.5**
f1_H2O_exp	**0**
A2	**4.50x10^{-2}**
E2, kJ/kgmole	**-9,395**
f2_H2O_exp	**1**
n, Denominator Exponent	**4**
α, (1/m^3)	**0.069198**
C_{pa}, kJ/kmol K	**70**
C_{pb}, kJ/kmol K	**110**
C_{pc}, kJ/kmol K	**37**

Initial reactor sizing included iterations with inlet temperature and concentrations to determine the effects of each on conversion and catalyst mass. It was determined that conversions greater than 70% could be achieved at high pressure and temperature, whereas conversions around 30% were achieved at more modest conditions. An Aspen HYSYS simulation was configured with a conversion reactor to determine the effects of conversion on sizing the remainder of the system. Through a number of iterations, it was seen that increases in conversion led to dramatically reduced column and heat exchanger sizes and utility costs. With this understanding, it was determined to design for the highest conversion at which the reactors can be safely and consistently be operated.

Through reaction simulation iterations on Polymath, the highest conversion that could be safely and consistently controlled was found to be approximately 70%. As the maximum temperature for the catalyst to operate at is 400°C, the process was designed to produce vapor with an outlet temperature between 375°C and 385°C. This range provides a reasonable buffer zone of 15°C between design outlet temperature and catalyst decomposition while maximizing the increased rate of reaction found at higher temperatures. While the increase in temperature leads to a decrease in gas concentration at a given pressure, the pressure needed to be increased to ensure sufficient gas concentration at the inlet to the reactor. A minimum inlet pressure was set to be 150 psia and was subsequently increased to maintain high conversion with reduced catalyst.

For reactor optimization the Aspen HYSYS simulation of the system provided the inlet flow rates and compositions for the reaction. These values were then exported to Polymath to determine the conversion and costing parameters. Once the conversion was set the reaction in HYSYS was altered to match the DE solver results. Changes in conversion altered the flow rates and composition to the reactors, and these were subsequently entered back into Polymath to find the new process output.

Using this process, the inlet pressure was increased until a reasonable maximum operating pressure was obtained. Each increase in pressure led to a decrease in catalyst required to achieve 70% conversion, but these pressure increases included process tradeoffs. Though the amount of catalyst and thus the size of the reactors decreased, the pressure factor for the thickness of the walls increased. More importantly, the increase in pressure must be accounted for in design of the heating and cooling of the feed and products. In addition, any increase in pressure must be produced by the methanol feed pumps, which would have an effect on overall capital and utility costs throughout the process. Finally, high-pressure systems can pose a hazard to health and safety, so it was decided to not exceed the minimum functional pressure for the reactor. After these considerations and numerous reaction simulations, 500 psia was taken to be the design pressure.

Once the design pressure was settled, the inlet temperature to the reactor was varied to find the value that produced an outlet temperature between 375 and 385°C (707-725°F). At an inlet temperature of 280°C (536°F), the discharge temperature was 382.7°C (721°F), which maximizes the kinetic advantages of high-temperature operation while mitigating the risks of catalyst decomposition. Final Polymath simulation parameters are displayed in Table 10. Selected Polymath data is shown below, additional charts and data can be found in Appendix A.

Table 10: Polymath Reaction Simulation Results

Parameter	Value
F_{ao}, kmol/s	**0.103**
C_{ao}, kmol/m^3	**0.7347**
C_{bo}, kmol/m^3	**0.00022**
C_{co}, kmol/m^3	**0.01469**
T_o, °F	**536**
Wfinal, ft3	**64**
Outlet pressure, psig	**465**
Outlet temperature, °F	**721**
Mass of catalyst per reactor per year, lb	**3,520**

As seen in Figure 4, the conversion of methanol and product concentrations exhibit a gradual increase across the length of the catalyst, while the pressure ratio and methanol concentrations decrease with increasing W. These charts show that there is continued advantage of additional catalyst, but the outlet temperature requirement and the increased cost over 20 years of catalyst replacement reduces the viability of extending the catalyst volume past two cubic meters. The design parameters of this system have been optimized to ensure both economic feasibility while maintaining ease of operational use and process safety.

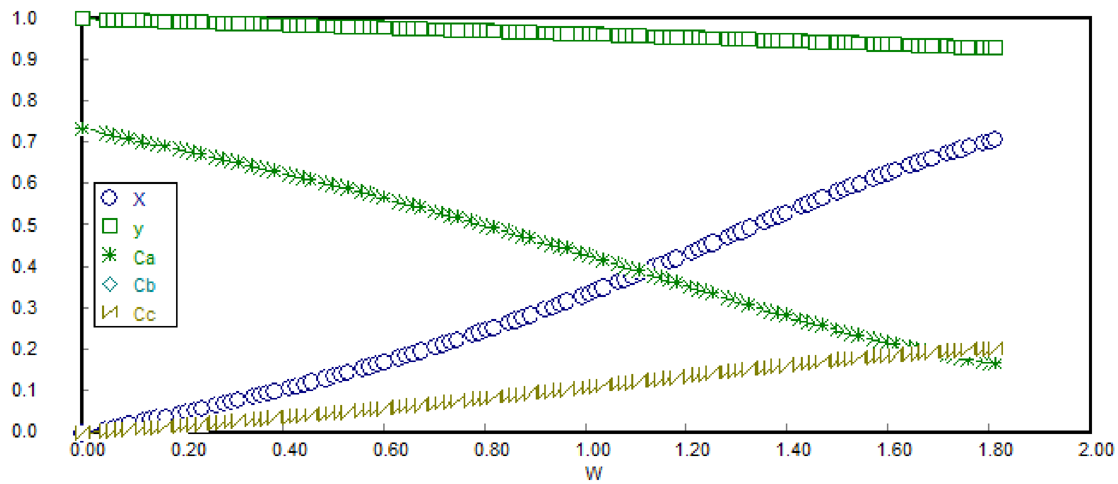


Figure 4: Plot of Conversion (X), Pressure Ratio (y), and Component Concentrations (kmol/m^3) vs. Catalyst Volume (W, m^3)

According to the stated design heuristics, the reactors were assumed to be vertical tanks packed with an array of parallel tubes loaded with catalyst (21). The tubes were assumed to be schedule 40 nominal piping, and the inner diameter of the vessel was set at 36 inches. Utilizing the sizing tool on Engineering ToolBox (23) to calculate the maximum number of smaller pipes in a larger pipe, a total of 80 tubes were sized to fit within the shell. With this design, four parallel packed-bed reactors will operate in parallel and produce a conversion of 70% using 3,520 pounds of ɣ-Al_2O_3. The reactors will be vertical pressure vessels designed to withstand 550 psia and 750°F. They are to be constructed out of carbon steel with a diameter of 3 feet and a height of 16 feet.

Heat Exchangers

The heat exchangers were designed to have two exchangers operate in parallel with one another. Not only does this ensure operation at the design pressure and temperatures at 50% turndown, but it also leads to reduced capital costs and lower cooling water expenditure. All HEX calculations were based on a HYSYS simulation of a 50% flow scenario and were doubled to meet total design capacity.

As the reactant feed stream required vaporization and the outlet gases needed to condense, the HYSYS simulation was set up with three heaters in series before the reactor and three coolers in series after the reactor. The first and third coolers/heaters were specified to produce sensible heat transfer, while the heat exchangers in between them were specified for latent heat transfer. With the heat transfer required and temperatures for each step known, a T-Q diagram was constructed and is displayed in Figure 5.

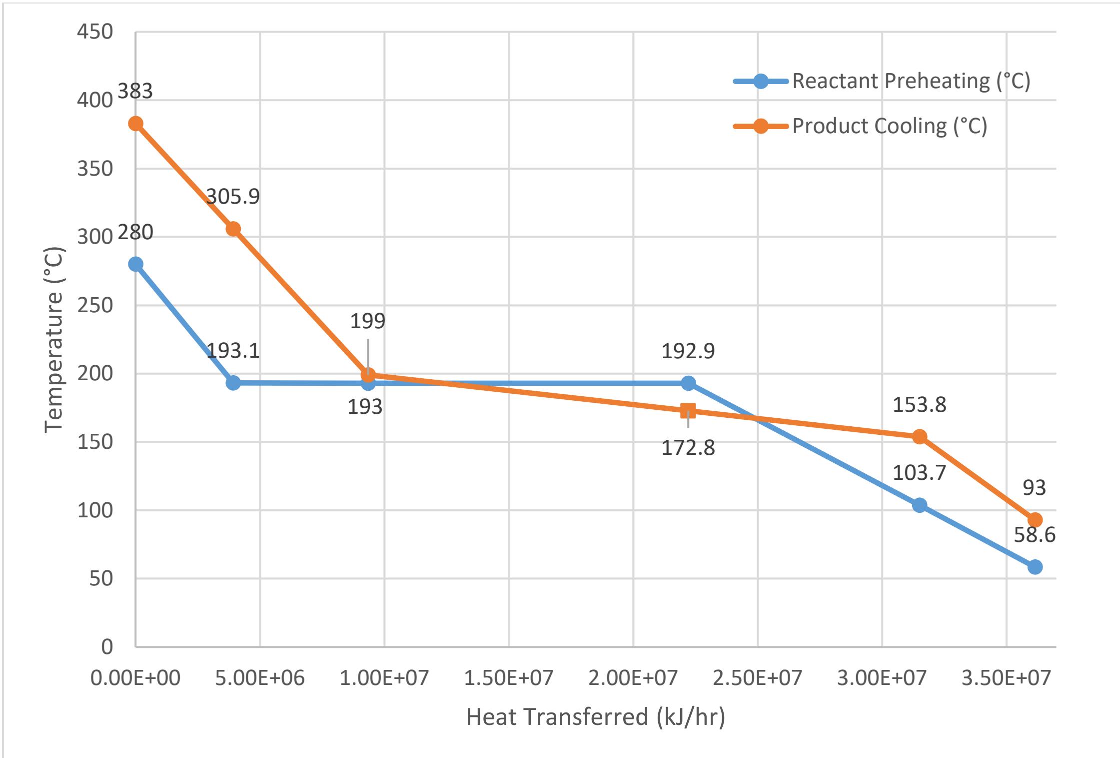


Figure 5: T-Q Chart for Heat Integration

There is a significant temperature cross that negates the possibility of total stream integration. Variations in the outlet vapor pressure were made, but it was not possible to achieve sufficient driving force for heat integration while maintaining the design parameters determined in reactor optimization. The primary challenge to the heat integration is that the outlet vapor mixture condenses along the range from 199°C to 153.8°C (390-309°F) while the essentially pure methanol fed to the reactor vaporized within a 0.2°C range.

A number of solutions were proposed to overcome this obstacle. Steam heating the inlet was considered, but the steam temperature was below the vaporization temperature of methanol at the given pressure. Another solution was to vaporize the methanol at a lower temperature and pressure and subsequently compress the vapor to the required pressure, but this was far too costly to be considered. The best process was determined to be using a furnace to vaporize the methanol and heat integration for sensible heat transfer. The remainder of the heat in the product stream is removed with cooling water. Figure 6 shows the revised heat transfer design.

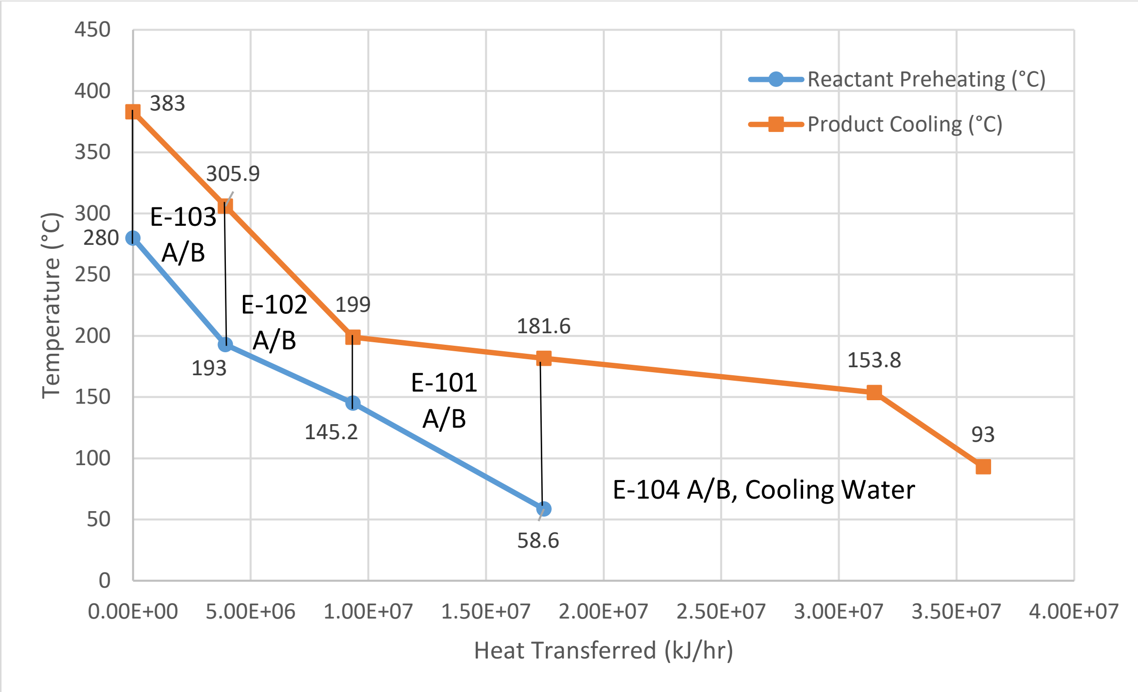


Figure 6: Updated heat integration design. Reactant preheating stream will be vaporized in the preheating furnace before the final sensible heat transfer is accomplished through heat integration.

The figure displays the design for the T-Q diagram. Liquid methanol will exit the pumps at 58.6°C (137°F) and absorb energy from the vapor cooling and condensing stream of product that enters as a saturated vapor at 199°C (390°F) and exits as a partially condensed fluid at 181.6°C (359°F). The methanol will exit this heat exchanger at 145.2°C (293°F) and enter E-102, where it will heat up until it becomes a saturated liquid at 193°C (379°F) and subsequently enter the preheating furnace. The product stream will enter the shell side of HEX-2 as a vapor and cool from 305.9°C (583°F) to a saturated vapor at 199°C (390°F) The furnace will vaporize the entire methanol feed, and the reactant stream will enter the second heat exchanger as a saturated vapor. The final sensible heat transfer will occur between the hot reactor vapor outlet entering at 383°C (721°F) and the methanol entering at 280°C (536°F). The partially condensed product vapor stream will be condensed and cooled to 93°C (199°F) using cooling water before entering the DME separation tower.

Each heat exchanger was sized according to the same process. Each portion of the heat exchangers where a difference in mode of heat transfer was detected was sized individually. The process is described using E-103 A/B, and data for the other exchangers sized can be found in Appendix B: HEX Calculations.

HEX Design Equation $$Q = U_o * A_o * F * LMTD \quad (14)$$

The first step in sizing each heat exchanger was to construct a T-Q diagram showing the heat transfer graphically as shown in figure 7. Each exchanger was set up in countercurrent configuration and the relative temperature differences at the outlet and inlet were calculated. From these values the log mean temperature difference (LMTD) was calculated. Table 11 and Figure 7 display the data for the vapor phase sensible heat transfer.

Table 11: Temperature Data for Vapor Phase Heat Integration

Q (kJ/h)	Reactant Preheating (°C)	Product Cooling (°C)	ΔT
0	280	383	103
3.92E+06	193.1	305.9	112.8
LMTD	$LMTD = \frac{(\Delta T_1 - \Delta T_2)}{ln(\frac{\Delta T_1}{\Delta T_2})}$		107.8

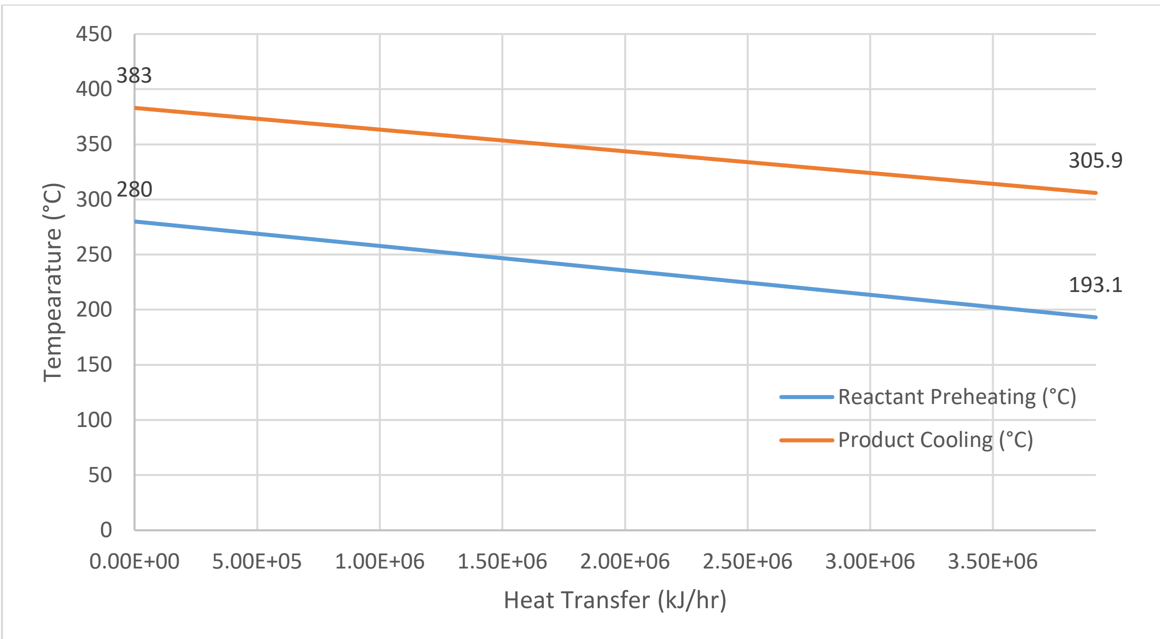


Figure 7: T-Q Diagram for Vapor Phase Heat Integration

The overall heat transfer coefficient U_o is calculated with Equation (15) from the CRC Handbook of Thermal Engineering (24).

$$U_o = \left[\frac{d_o}{h_i d_i} + R_{fi}\frac{d_o}{d_i} + \frac{d_o ln(\frac{d_o}{d_i})}{2K_w} + R_{fo} + \frac{1}{h_o}\right]^{-1} \qquad 15$$

The equation shown above requires knowing the inner and outer diameters of the heat exchanger tubes. As this is not known, the value of $\frac{d_o}{d_i}$ was assumed to be 1, and the decrease in convective heat transfer on the tube side was assumed to counteract the loss of resistance due to wall resistance. This assumption is justified by simplifying the calculations above significantly and by the result of the assumption providing a conservative estimate for the value of U_o. Erring on the conservative side for the heat transfer coefficient leads to larger values for heat transfer area, which will ensure sufficient sizing on the heat exchangers.

The convective heat transfer coefficients and fouling factors used in the analysis were taken from the CRC Handbook of Thermal Engineering (24). Table 4.1.1 provided ranges for both h and fouling resistance for different types of heat transfer with chemicals under varying conditions. The values taken from this table were assumed to be in the lower half of the practical range, which ensured that the heat transfer coefficient was always a conservative estimate.

The correction factor for the heat exchangers was found as a function of dimensionless temperature ratios P and R, where P and R are defined as follows:

$$P = \frac{T_{tube,out} - T_{tube,in}}{T_{shell,in} - T_{tube,in}} \quad (16)$$

$$R = \frac{T_{shell,in} - T_{shell,out}}{T_{tube,out} - T_{tube,in}} \quad (17)$$

P and R are then used to find F from the correction chart. The value determined for the vapor phase sensible heat exchanger was estimated to be 0.98. This value was checked using the heat transfer LMTD calculator at CHECalc.com (25). The HEX utilized in the analysis of these heat exchangers was assumed to be 2-shell pass, 4-tube pass, and the value obtained in this way was 0.9751. This value was utilized in all subsequent calculations.

With Q, U_o, F, and LMTD known, the area of the heat exchanger was solved for by rearranging the design equation. This area was then multiplied by 1.1, which serves as a sizing factor to ensure adequate surface area for non-idealities and process control measures. This area was utilized in costing to determine the approximate expenditure required to install the exchanger.

The cooling water heat exchanger was sized according to the process described above, but additionally required calculation of cooling water flow rates. The outlet temperature for the cooling water to return to the cooling towers was set at 120°F (49°C) and data sources set the heat capacity of water at 75.37 kJ/kmol*K. Assuming a constant C_p, the mass flow rate of water can be found using Equation 18 and was converted to find a water flow rate of 1,040 gpm.

$$\dot{m} = \frac{Q}{C_p \Delta T} \quad (18)$$

In contrast to the heat exchangers, which were designed to have two exchangers operate in parallel to meet design capacity, one furnace was designed to vaporize the entire feed of the methanol to the reactor. As furnaces are sized purely on the energy that they are required to deliver, the energy required to vaporize the total methanol feed stream as determined by HYSYS was converted to kW and used to cost the unit. No additional calculations were undertaken for sizing.

Each of the heat exchangers utilized in this system are AEL TEMA type heat exchangers made out of carbon steel and designed for a pressure of 580 psia. The AEL TEMA type is a widely used for a multitude of applications in the industry and provides flexibility in both condensation and evaporation as well as sensible heat transfer (24). A summary of the overall heat integration strategy is displayed in Table 12.

Table 12: Summary of HEX Design Functions

Heat Exchanger Number	Function	Tube Side Stream	Shell Side Stream
E-103 A/B	**Vapor Phase Sensible Heat Transfer**	**Reactant Preheating**	**Product Cooling**
E-102 A/B	**Liquid Heating and Vapor Cooling**	**Reactant Preheating**	**Product Cooling**
E-101 A/B	**Liquid Heating and Vapor Condensing**	**Reactant Preheating**	**Product Cooling**
E-104 A/B	**Condensing and Vapor Cooling**	**Cooling Water**	**Product Cooling**
E-203 A/B	**Wastewater Cooling**	**Cooling Water**	**Wastewater**
H-101	**Reactant Vaporization**	**n/a**	**n/a**

While this strategy is the most cost-efficient method that can be reasonably implemented, there is one drawback. The vapor product in E-102 enters as a superheated vapor but exits as a two-phase mixture. It is never ideal to have two-phase flow because it is unsteady and leads to increased corrosion and wear on the system. In this case, however, the economic benefits of operating the heat exchangers in the order that they are implemented justifies the possibility of increased maintenance on the line between the shell side of E-101 and E-104. In the plant, these could be combined into a multi-shell, multi-tube pass system in which the first tube passes are the methanol feed and the remaining tube passes are cooling water. This would eliminate the negative effects of two-phase flow through an extended length of pipe.

T-101 A/B

T-101 A/B is two distillation columns acting in parallel with each column producing half of the desired DME per day. In order to design T-101 A/B to produce our specified DME composition, 99.95% by mass, the Fenske equation (19) was used to determine the minimum number of stages and the Underwood equation (20) was used to determine the minimum reflux ratio. Using these equations, it was found that the minimum number of stages was 7 and the minimum reflux ratio was 0.212.

$$\mathrm{Nmin} = \frac{ln\left(\frac{x_D(1-x_B)}{x_B(1-x_D)}\right)}{ln(\alpha_{AB})} \qquad 19$$

$$R_{min} + 1 = \sum_{i=1}^{c} \frac{\alpha_{i,y} x_D}{\alpha_{i,y} - \Phi} \qquad 20$$

The efficiency of towers, T-101 A/B, was calculated using the Drickamer-Bradford Equation due to the hydrocarbon nature of the mixture, which can be seen in Equation 21. For the Drickamer-Bradford equation to calculate the efficiency, T-101 A/B had to be operating within the following parameters: average temperature between 157-420 °F, pressure between 14.7-366 psia, liquid viscosity (μ) between 0.066-0.355 cP, and an overall efficiency (E_o) between 41-88 %. Table 13 shows that the efficiency of T-101 A/B to be 77% (26).

$$EO = 13.3 - 66.8log(\mu) \quad 21$$

Table 13: Drickamer-Bradford Equation

Average Liquid Viscosity, μ	**0.113**
Efficiency %, EO	**77**

After determining the efficiency of T-101, the most cost efficient tower was designed by first plotting reflux ratio versus the number of stages as shown in Figure 8. The chart of N vs. R shows that the most cost efficient design is likely 23-30 actual trays. Multiple towers were costed in the 23-30 actual tray range in an attempt to find the optimum number of trays. Figure X below suggests that T-X is most cost efficient at 27 actual trays.

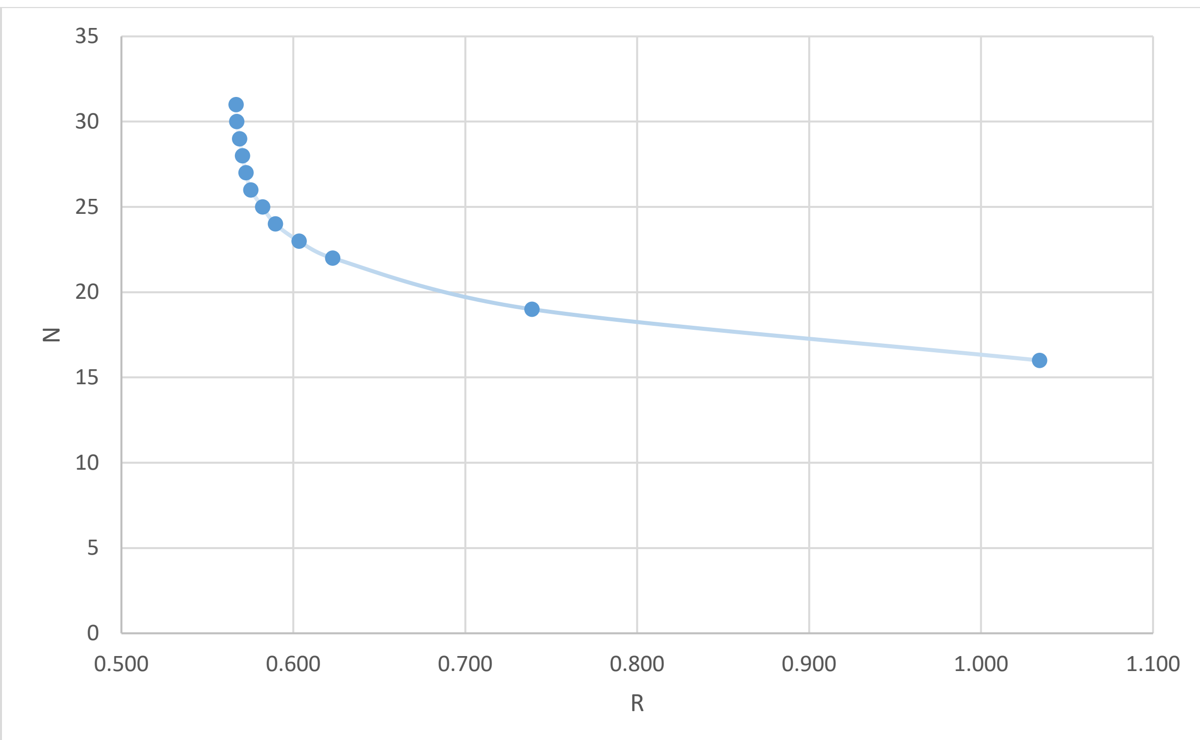


Figure 8: A graph of N vs R

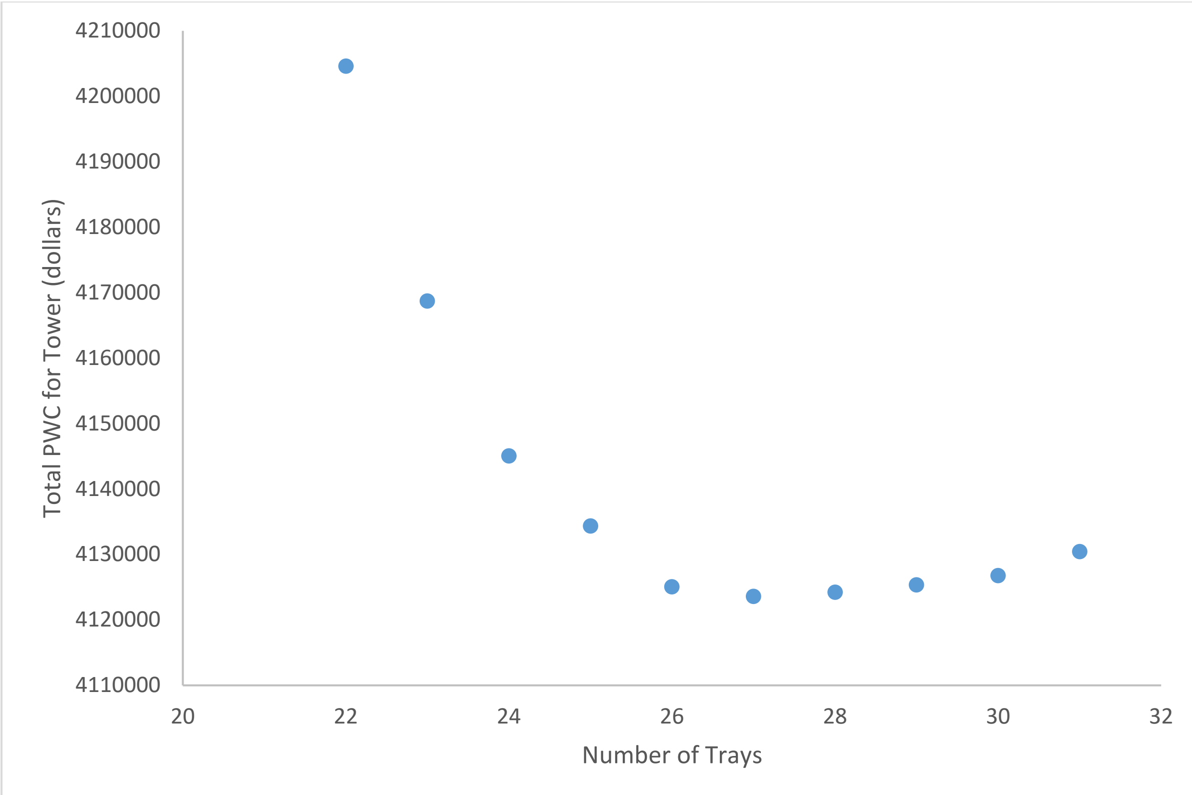


Figure 9: Total cost of tower versus number of actual trays

The column's internals and feed stage were optimized by plotting total PWC of the tower vs. number of trays (Figure 9). The optimum number of trays was determined to be 27. The optimum feed stage was defined as the stage at which the reflux ratio was the lowest as this minimizes the cost of the condenser (E-105 A/B) and reboiler (E-106 A/B) utility streams. Through iterative simulations the optimum feed stage was found to be at stage 8. The results of these iterations can be seen below in Table 14 and Figure 10.

Table 14: Reflux ratio at various feed stages for the 27 stage design

Number of Trays	Feed Stage	Reflux Ratio
27	7	0.5945
27	8	0.5757
27	9	0.5723
27	10	0.5751
27	11	0.5823

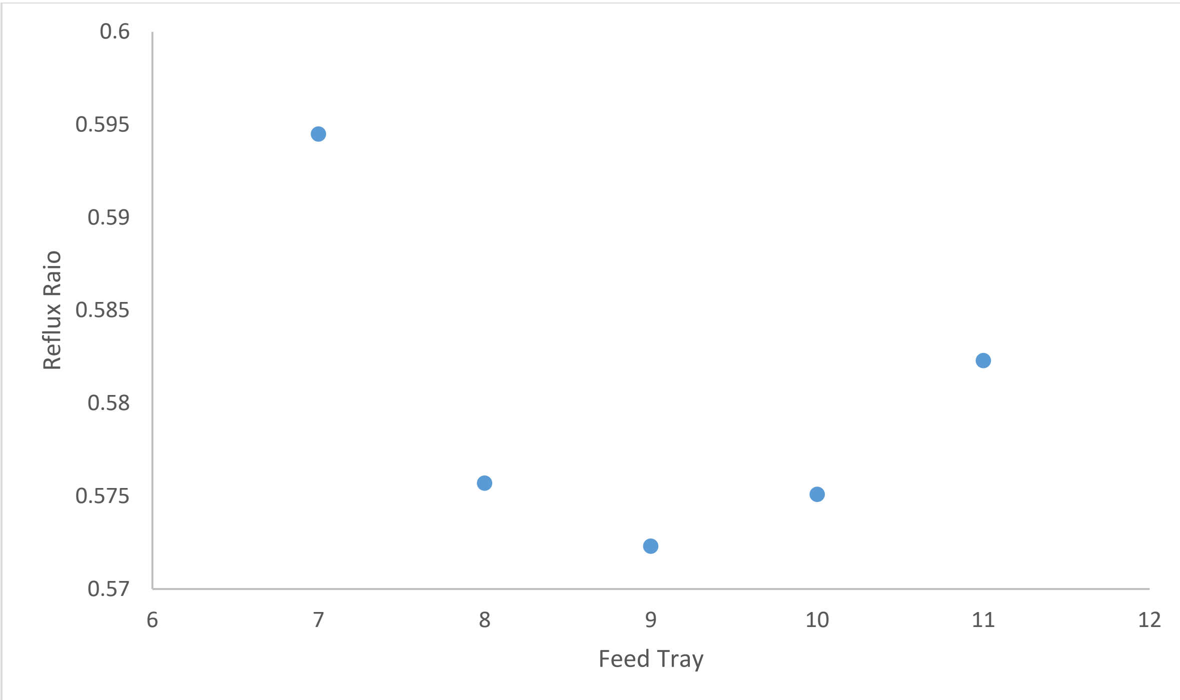


Figure 10: A plot of reflux ratio versus feed tray for the 27 actual tray simulation

Heuristics from Turton prescribe that tray spacing be between 18 and 24 inches (21). Since the total height of the column was not a major concern in the design, 24 inch tray spacing was used to allow more room for maintenance. With the tray spacing set at 24 inches, the diameter was then varied in an attempt to minimize the cost of the column while keeping an approach to flood below 80%. At 24-inch tray spacing it was found that 3 feet diameter trays minimized the cost of the column and had an approach to flood of 79%. The team used sieve trays for the design as they are the most cost efficient tray type for this process. Table 15, Table 16, Table 17, and Table 18 below show the parameters for the column including tray, condenser and reboiler parameters.

Table 15: T-101 A/B Parameters

Select Carbon Steel			
Parameter	**Equation/Source**	**Value**	**Units**
Volume: Area*Height (ft3)	Volume: Area*Height (ft3)	388.58	ft3
Height	Column Height= htop+htrays+hbtms	55	ft
htop	Tray Spacing plus 1 ft	2	ft
htrays	From Hysys	46	ft
hbtms	(3 ft for vapor/liquid disengagement and 4 ft for holdup)	7	ft
Area	Area= (pi/4)* (Diameter)^2	7.07	ft2
Diameter	From Hysys	3	ft
Pressure	Converting psia to barg	14.3	barg
Number of theoretical stages	Original HYSYS output	18	stages
Number of actual trays	Number of actual trays=(number of theoretical trays/Efficiency)	23	trays
Efficiency	Drickamer-Bradford	77	%

Table 16: T-101 A/B Trays

Sieve Trays			
Parameter	Source	Value	Units
Area, A	Area (ft2) from HYSYS	7.065	ft2
Diameter	From HYSYS	3	ft

Table 17: Condenser E-105 A/B

Parameter	Value	Units
Area	1598	ft2
Area with safety factor	1757.8	ft2
LMTD	14	N/A
U	850	W/m2K
Duty	7.60*106	kJ/hr
Pressure	14.1	Barg
Temperature	47.37	C

Table 18: Reboiler E-106 A/B

Parameter	Value	Units
Duty	8.81 * 106	(KJ/hr)
Temperature	151.7	C
Steam Pressure	41	barg
Steam Temperature	254	C
Latent Heat	2019	KJ/Kg
LMTD	102.3	N/A
U	850	(W/m2K)
F	0.9751	N/A
A	310.7	ft2
A with Safety Factor	341.8	ft2
Pressure	14.3	barg

T-201 A/B

Using the same procedures as in T-101 A/B, a second distillation column was designed to remove wastewater from the process and recycle the unreacted methanol. T-202 A/B is two distillation columns working in parallel each removing half of the wastewater from the system. It was determined that the minimum number of stages is 8 and the minimum reflux ratio is 1.111. All calculations and tables are in Appendix C: Tower Calculations.

Tower-201 A/B contains a binary mixture; therefore, the O'Connell Correlation as seen in Equation 22 was used for calculating tray efficiency as opposed to the Drickamer-Bradford equation. Using the O'Connell equation, the efficiency of T-201 A/B was found to be 72% as shown in Table 19.

$$E_O = 50.3\ (\alpha^*\mu)^{-.226} \qquad 22$$

Table 19: O'Connell Correlations

O'Connell Correlation	
α	2.93
μ	0.0688
Eo %	72

Pumps:

The pumps were designed to pump the liquid streams to a desired pressure to flow through the system. The pump types were selected using Figure 11, which uses the specific head in feet and flow rate in gpm for each individual pump. Some pump selections have been shown in Figure 11. The pumps were sized based on hand calculations of hydraulic power and break horse power. Using Equation 23, the head, flow rate, specific gravity of the fluid, and a safety factor were required for finding the hydraulic power in kilowatts. A safety factor of 10 % was used in the calculations. The brake horsepower was then calculated by using the efficiency of the pump and Equation 26. Efficiencies were found using the pump heuristics Table 11.9 in the Turton design book (21). All values for actual flow rate and specific gravity were taken directly from the HYSYS simulation. For every pump throughout the system, another pump was purchased as a spare. The spares are installed in parallel in the even that either pump stop working there will be a spare to continue operations. All pumps were using the same method; complete pump calculations are shown in Appendix D: Pump Sizing Calculations.

P-101 A/B/C/D

The first pump in the process was the methanol feed pressurizing pump, P-101 A/B. The pumps were designed to pressurize the methanol feed to a desired pressure of 530 psia. The pressurized methanol would then flow into multiple preheaters and then catalytic reactor to be converted into DME. Using the required outlet pressure and the inlet pressure of the pump, the differential pressure across the pump was calculated to be 485.4 psia. Using equation (23) the hydraulic power was calculated to be 33.20 kW. Using Figure 11, a multistage centrifugal pump was selected based on its specific head and flow rate. After selecting the pump type, the efficiency was estimated to be 50 % using heuristics found in Turton (21). The brake horsepower was calculated to be 66.40 kilowatts. Carbon steel was selected as the material of construction for P-101 A/B, as methanol would corrode a cast iron pump. The spares pumps, P-101 C/D have been installed in parallel in order to continue production if P-101 A/B were to fail.

Table 20: Pump Calculation Summary for P-101

P - 101 A/B/C/D		
Parameter	Value	Units
Flow Rate:	141	gal/min
Head:	1514	ft
Pump Selection	Multi-Stage Centrifugal	
Pressure	500	psia
Hydraulic hp	33.2	kW
Efficiency	50	%
Break hp	66.4	kW

Hydraulic Power (kW) 23

$$HP = \frac{Q \, * \, \Delta P \, * Safety\ Factor}{1715 \, * 1..34}$$

Head (ft) 24

$$Head = \frac{2.31 * \Delta P}{Specific\ Gravity}$$

ΔP (psi) 25

$$\Delta P = \frac{Head\ (ft) * Specific\ Gravity}{2.31}$$

Brake Horse Power (kW) 26

$$\text{BHP} = \frac{\text{Hydraulic Power}}{\text{Efficiency of Pump}}$$

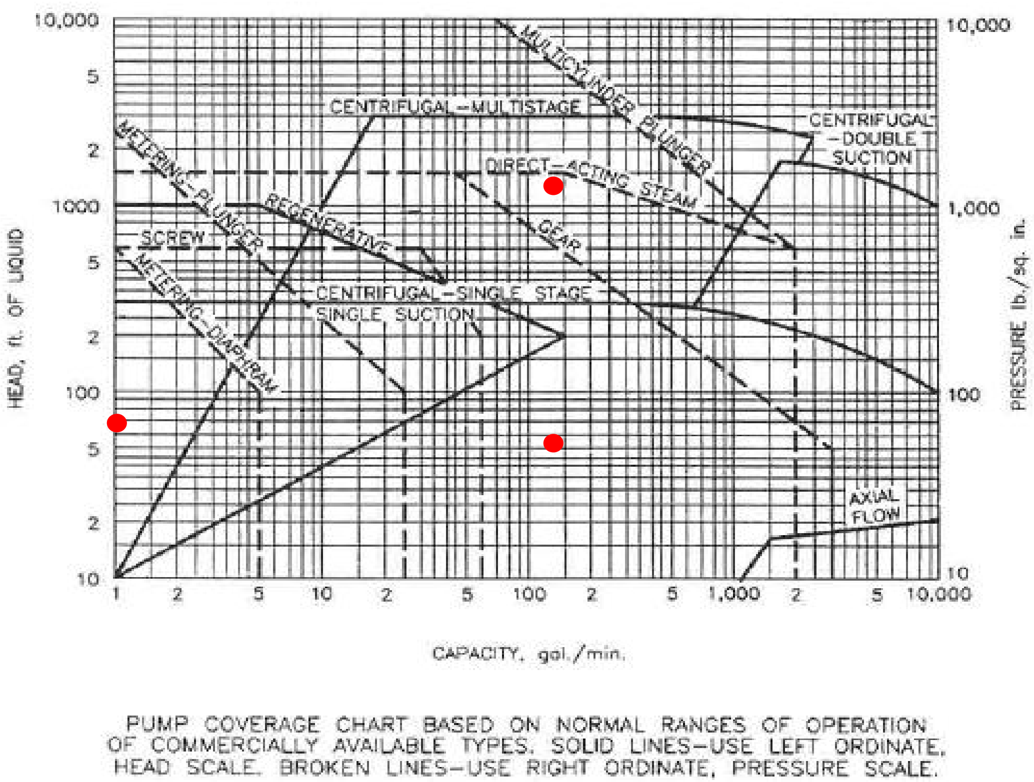


Figure 10-14. Pump selection guide (*from Gas Processors Suppliers Association*).

Figure 11: Pump Selection Chart from GPSA Handbook

Vessel Sizing:

The condensate receivers were designed to temporarily store the liquid product coming out of the condensers for T-101 A/B and T-201 A/B, to provide Net Positive Suction Head (NPSH) for the reflux pump, and to protect against any surges. Using the pressure vessel sizing heuristics from Turton, the vessels were designed to have a volume equivalent of 10 minutes of hold up. The vessels were also designed to have an L/D factor of 3 with this value being the optimum ratio that Turton heuristics suggests.

The first condensate receivers, V-101 A/B, were designed to have 200 ft^3 of storage. The volume of the receiver was calculated from Equation 27 using the flow rate entering from the condenser of 1200ft^3/s and the 10 minutes of liquid holdup. Next, the diameter of the vessel was calculated from Equation 29 and then using the optimum ratio from Equation 28. The diameter of the vessel was then calculated to be 4.4 ft and the length was found to be 13.2 ft.

The second condensate receivers, V-201 A/B, were designed the same way as V-101 A/B. Using values from HYSYS for the flow entering the receiver of 767.4 ft^3/s and the 10 minutes of holdup, a volume of 127.9 ft^3 was calculated. Using Equations 28 and 29, a diameter of 3.8 ft. was calculated. With the calculated diameter, the required length of the vessel was found to be 11.4 ft.

Volume (ft^3)	$Q * t$	27
Length (ft)	$L = 3 * D$	28
Diameter (ft)	$\sqrt[3]{\frac{4V}{3\pi}}$	29

Tk-101 A/B

With such a large production rate of DME, storage tanks will be needed to maintain the capacity of methanol feed required for the process. Through process simulation, the daily volume of methanol needed was found to be approximately $2.72*10^5$ gallons/day. Based on the storage tank heuristics given in Turton (21), the capacity of the storage tanks were designed to hold a 30-day supply of raw material. Volumes for each tank can be seen in Table 21. Another heuristic for the storage tanks assumed a freeboard space of 10% for tanks larger than 500 gallons. Since these methanol tanks will be significantly larger than that specification, a freeboard and safety factor design volume of 20% was added onto the required liquid volume for the tanks. Through size iterations and cost optimization, the number of storage tanks needed for the plant was determined to be two tanks with size parameters of 138 ft. diameter and 44 ft. height for each tank. Based on the diameter of the tanks, floating roof tanks were chosen as the type of tank since they are more practical in constructing large diameter tanks and are more cost efficient for the tank volume needed.

Table 21: Parameters for Storage Tank Sizing

Parameter	Value
Daily Volume (gal/hr)	**11,322**
Actual Storage Volume needed (gal/30day)	**8,151,840**
Design Volume (gal/30day)	**8,152,000**
Number of Tanks	**2**
Liquid level volume needed (gal/tank)	**4,076,000**
Design Volume with 20% Safety Factor (gal/tank)	**4,891,200**
Diameter (ft)	**138**
Height (ft)	**44**
Volume based on dimmensions (gal/tank)	**4,923,027**

Since methanol is a polar and conductive compound, corrosion on the inner walls poses a large threat to the life cycle of the storage tank walls (27). Both carbon steel and stainless steel were considered in the design of the tanks. While carbon steel offers a lower capital cost investment, there are unpredictable corrosion based maintenance costs associated with using carbon steel. Stainless steel has a capital cost approximately three times that of carbon steel, but offers a more secure, low maintenance risk concerning corrosion effects on the inner walls of the tanks. This design will use stainless steel since the process is very dependent on a pure methanol feed and cannot risk contaminating the feed tank with methanol that has reacted with other compounds of the carbon steel.

Mixer and lubricant tank

With DME having a poor lubricity property, lubrication is necessary before shipping the product off for usage as a fuel (28). Although there is no definitive specification for the amount of lubricant needed in DME fuel, this plant will consider 900 ppm by mass of lubricant sufficient for the DME fuel. In order to add in the lubricant, a storage tank, Tk-301, and mixer, M-302, were put in place before the DME is sent to the truck filling station. Much like the storage tanks for methanol, the lubricant tank was designed to hold a 30-day capacity of lubricant needed for production. In order to design the storage tank, the amount of lubricant needed on a daily basis and then multiplied for 30 days. The monthly amount of lubricant needed was found to be 4,600 gallons. Using Turton (21) heuristics, a freeboard volume of 10% was added in addition to the lubricant volume. Using the heuristic of an L/D value equal to 3, the length and diameter of the tank were found to be 19.2 ft. and 6.4 ft. respectively. Although the tank will be operating at atmospheric pressure, the design pressure was determined to be 50 psi above the operating pressure based on heuristics for vessels given in Turton (21). Lastly, the vessel was designed to be made of carbon steel since it is provides a smaller capital investment.

The mixer will combine the two DME production streams from towers, T-101 A/B, and the lubricant stream to ensure that the DME produced will be lubricated enough to protect vehicle engines. To meet the production demand of DME, a horizontal mixer was designed capable of handling flowrates of approximately 11,000 gal/hr. Using Turton heuristics, a holdup of 10 minutes was used in sizing the mixer, and an L/D ratio 3 was assumed for size parameters when determining volume of the mixer vessel. Optimizing the diameter and length of the mixer resulted in a diameter of 4.7 ft. and length of 14.1 ft. With these size parameters the design volume of the mixer was found to be 1,830 gallons. Along with designing the size of the mixer, the impeller power requirements were also determined. Using an estimate for the power requirement of 0.3 kW/m^3, a power requirement for the impeller was calculated to be 2.08 kW.

Equipment Specification Sheets

Reactor Specifications

Unit:	Reactors R-1, A/B/C/D
Type & Sizing	
Orientation	Vertical
Internal Diameter	3 feet
Length	16 feet
Head Type	2:1 elliptical
Number of Internal Tubes	80
Nominal Size of Internal Tubes	3"
Conditions	
Fluid in	Methanol
Fluid out	Methanol, DME, Water mixture
Specific Gravity, Feed	1.11
Operating Temperature	720°F
Design Temperature	750°F
Operating Pressure	500 psia
Design Pressure	550 psia
Nozzles/Connections	
Connection	Service
1	Reactor Inlet
2	Reactor Outlet
3	Pressure Relief

E-101 Specifications

Unit:	E-101 A/B
Type & Sizing	
Orientation	Horizontal
Heat Transfer Area	502 square feet
TEMA Type	AEL
Head Type	Fixed Tube
Function	Vapor condensation, liquid phase heating
Conditions	
Shell Fluid	Methanol
Tube Fluid	Methanol, DME, Water mixture
Max Operating Temperature	720°F
Design Temperature	750°F
Operating Pressure	524 psia
Design Pressure	580 psia
Temperature, Shell Side In	390°F
Temperature, Shell Side Out	359°F
Temperature, Tube Side In	137°F
Temperature, Tube Side Out	293°F
Nozzles/Connections	
Connection	Service
1	Cool Vapor In
2	Heated Vapor Out
3	Hot Product Vapor In
4	Cooled Product Vapor Out

E-102 Specifications

Unit:	E-102 A/B
Type & Sizing	
Orientation	Horizontal
Heat Transfer Area	552 square feet
TEMA Type	AEL
Head Type	Fixed Tube
Function	Vapor phase cooling, liquid phase heating
Conditions	
Shell Fluid	Methanol
Tube Fluid	Methanol, DME, Water mixture
Max Operating Temperature	720°F
Design Temperature	750°F
Operating Pressure	516 psia
Design Pressure	580 psia
Temperature, Shell Side In	583°F
Temperature, Shell Side Out	390°F
Temperature, Tube Side In	293°F
Temperature, Tube Side Out	380°F
Nozzles/Connections	
Connection	Service
1	Cool Vapor In
2	Heated Vapor Out
3	Hot Product Vapor In
4	Cooled Product Vapor Out

E-103 A/B Specifications

Unit:	E-103 A/B A/B
Type & Sizing	
Orientation	Horizontal
Heat Transfer Area	402 square feet
TEMA Type	AEL
Head Type	Fixed Tube
Function	Vapor phase sensible heat exchange
Conditions	
Shell Fluid	Methanol
Tube Fluid	Methanol, DME, Water mixture
Max Operating Temperature	720°F
Design Temperature	750°F
Operating Pressure	508 psia
Design Pressure	580 psia
Temperature, Shell Side In	721°F
Temperature, Shell Side Out	583°F
Temperature, Tube Side In	380°F
Temperature, Tube Side Out	536°F
Nozzles/Connections	
Connection	Service
1	Cool Vapor In
2	Heated Vapor Out
3	Hot Product Vapor In
4	Cooled Product Vapor Out

E-104 Specifications

Unit:		E-104 A/B
Type & Sizing		
Orientation		Horizontal
Heat Transfer Area		502 square feet
TEMA Type		AEL
Head Type		Fixed Tube
Function		Vapor condensation, liquid phase heating
Conditions		
Shell Fluid		Methanol
Tube Fluid		Methanol, DME, Water mixture
Max Operating Temperature		720°F
Design Temperature		750°F
Operating Pressure		524 psia
Design Pressure		580 psia
Temperature, Shell Side In		359°F
Temperature, Shell Side Out		199°F
Temperature, Tube Side In		87°F
Temperature, Tube Side Out		120°F
Nozzles/Connections		
Connection		Service
	1	Two-phase Product In
	2	Cooled Liquid Product Out
	3	Cooling Water In
	4	Cooling Water Out

H-101 Specification Sheet

Unit:	H-101
Type & Sizing	
Material	Carbon Steel
Energy Transferred	3.62e7 Btu/hr
Function	Liquid Vaporization
Conditions	
Process Stream	Methanol
Fuel	Fuel gas
Process Stream Outlet Temperature	380°F
Process Stream Inlet Temperature	379°F
Operating Pressure	516 psia
Design Pressure	580 psia
Nozzles/Connections	
Connection	Service
1	Liquid Reactant Stream In
2	Vapor Reactant Stream Out
3	Fuel Gas Feed Stream
4	Air Inlet

Pump Specifications

Unit:	P – 101, A/B/C/D
Type & Sizing	
Pump Type	Multistage Centrifugal
Hydraulic Power	33.20 kW
Brake Horse Power	66.40 kW
Function	Pressurizing Methanol Feed
Conditions	
Pump Fluid	Methanol, Water, DME
Head	1515 ft
Actual Flow Rate	140.8 gpm
Efficiency	50 %
Pressure Across Pump	485 psi
Operating Outlet Pressure	500 psia
Design Outlet Pressure	550 psia
Nozzles/Connections	
Connection	Service
1	Non-pressurized Methanol Feed In
2	Pressurized Methanol Feed Out

Unit:	P – 102, A/B/C/D
Type & Sizing	
Pump Type	Single Stage Centrifugal
Hydraulic Power	1.11 kW
Brake Horse Power	2.22 kW
Function	Pumping Reflux to Tower and DME to Mixer
Conditions	
Pump Fluid	Methanol, Water, DME
Head	64 ft
Actual Flow Rate	147 gpm
Efficiency	50 %
Pressure Across Pump	15.8 psi
Operating Outlet Pressure	169.5 psia
Design Outlet Pressure	219.5 psia
Nozzles/Connections	
Connection	Service
1	Condenser E – 105 Product In
2	Pumped Liquid Out

Unit:	P – 103, A/B/C/D
Type & Sizing	
Pump Type	Single Stage Centrifugal
Hydraulic Power	0.29 kW
Brake Horse Power	0.82 kW
Function	Feeding Bottoms Product into T-102
Conditions	
Pump Fluid	Methanol, Water, DME
Head	26.25 ft
Actual Flow Rate	71.28 gpm
Efficiency	35 %
Pressure Across Pump	8.4 psi
Operating Outlet Pressure	164.8 psia
Design Outlet Pressure	214.8 psia
Nozzles/Connections	
Connection	Service
1	T – 101 Bottoms Product In
2	Pumped Product Out

Unit:	P – 201, A/B/C/D
Type & Sizing	
Pump Type	Single Stage Centrifugal
Hydraulic Power	1.16 kW
Brake Horse Power	2.32 kW
Function	Pumping reflux into T 201 and distillate back into process as recycle
Conditions	
Pump Fluid	Methanol, Water, DME
Head	88.75 ft
Actual Flow Rate	94.57 gpm
Efficiency	50 %
Pressure Across Pump	25.6 psi
Operating Outlet Pressure	150.3 psia
Design Outlet Pressure	200.3 psia
Nozzles/Connections	
Connection	Service
1	Condenser E- 201 Product In
2	Pumped Liquid Out

Unit:	P – 301, A/B
Type & Sizing	
Pump Type	Positive Displacement Metering Pump
Hydraulic Power	0.003 kW
Brake Horse Power	0.009 kW
Function	Pumping lubricant from storage to mixer
Conditions	
Pump Fluid	Lubricant
Head	45 ft
Actual Flow Rate	0.11 gpm
Efficiency	35 %
Pressure Across Pump	60.3 psi
Operating Outlet Pressure	75 psia
Design Outlet Pressure	125 psia
Nozzles/Connections	
Connection	Service
1	Lubrication In
2	Pumped Lubrication Out

Unit:	P – 202, A/B
Type & Sizing	
Pump Type	Single Stage Centrifugal Pump
Hydraulic Power	0.15 kW
Brake Horse Power	0.44 kW
Function	Pumping Wastewater to treatment
Conditions	
Pump Fluid	Water and Methanol
Head	45 ft
Actual Flow Rate	24.0 gpm
Efficiency	35 %
Pressure Across Pump	10 psi
Operating Outlet Pressure	138.4 psia
Design Outlet Pressure	188.4 psia
Nozzles/Connections	
Connection	Service
1	Waste Water Product In
2	Pumped Waste Water Product Out

Vessel Specifications

Unit:	V – 101, A/B
Type & Sizing	
Tower Type	Condensate Receiver
Orientation	Horizontal
Volume	200.2 ft^3
Diameter	4.4 ft
Length	13.2 ft
Function	Store Condensate from T – 101 and Provide NPSH for Reflux Pump
Conditions	
Tower Fluid	DME/Water/Methanol
Design Pressure	203.7 psia
Operating Pressure	153.7 psia
Material of Construction	Carbon Steel
Nozzles/Connections	
Connection	Service
1	Condensate In
2	Condensate to Reflux/Distillate Out

Unit:	V – 201, A/B
Type & Sizing	
Tower Type	Condensate Receiver
Orientation	Horizontal
Volume	127.9 ft^3
Diameter	3.8 ft
Length	11.4 ft
Function	Store Condensate from T – 201 and Provide NPSH for Reflux Pump
Conditions	
Tower Fluid	DME/Water/Methanol
Design Pressure	174.7 psia
Operating Pressure	124.7 psia
Material of Construction	Carbon Steel
Nozzles/Connections	
Connection	Service
1	Condensate In
2	Condensate to Reflux/Distillate Out

Tower Specifications

Unit:	T – 101, A/B
Type & Sizing	
Tower Type	Distillation Column
Height	64 ft
Volume	388.6 ft^3
Diameter	3 ft
Tray Spacing	2 ft
Function	Separating DME and Methanol/Water
Conditions	
Tower Fluid	DME/Water/Methanol
Feed Stage from Top	9
Number of Actual Trays	27
Tray Efficiency	77 %
Tray Type	Sieve Trays
Design Pressure	206.4 psia
Operating Pressure	156.4 psia
Material of Construction	Carbon Steel
Nozzles/Connections	
Connection	Service
1	Feed Stream In
2	Vapor Product Out
3	Reflux Stream In
4	Liquid Product Out
5	Boil-Up Stream In

Unit:	T – 201, A/B
Type & Sizing	
Tower Type	Distillation Column
Height	88.75 ft
Volume	515.8 ft^3
Diameter	3 ft
Tray Spacing	1.75 ft
Function	Separating DME/Methanol and Water
Conditions	
Tower Fluid	DME/Water/Methanol
Feed Stage from Top	30
Number of Actual Trays	45
Tray Efficiency	72 %
Tray Type	Sieve Trays
Design Pressure	178.4 psia
Operating Pressure	128.4 psia
Material of Construction	Carbon Steel
Nozzles/Connections	
Connection	Service
1	Feed Stream In
2	Vapor Product Out
3	Reflux Stream In
4	Liquid Product Out
5	Boil-Up Stream In

Condenser Specifications

Unit:	E – 105, A/B
Type & Sizing	
Exchanger Type	Condenser
Area	2131 ft^2
TEMA Type	AEL
Head Type	Fixed Tube
Function	Condensing Vapor Leaving T - 101 for Reflux and Distillate Product
Conditions	
Shell Fluid	DME/Water/Methanol
Tube Fluid	Cooling Water
Max Operating Temperature	150 °F
Design Temperature	117 °F
Operating Pressure	169 psia
Design Pressure	219 psia
Nozzles/Connections	
Connection	Service
1	Cooling Water In
2	Cooling Water Out
3	Hot Process Stream In
4	Cooled Process Stream Out

Unit:	E – 201, A/B
Type & Sizing	
Exchanger Type	Condenser
Area	567 ft^2
TEMA Type	AEL
Head Type	Fixed Tube
Function	Condensing vapor leaving T - 101 for reflux and distillate product
Conditions	
Shell Fluid	DME/Water/Methanol
Tube Fluid	Cooling Water
Max Operating Temperature	300 ^oF
Design Temperature	269 ^oF
Operating Pressure	140 psia
Design Pressure	190 psia
Nozzles/Connections	
Connection	Service
1	Cooling Water In
2	Cooling Water Out
3	Hot Process Stream In
4	Cooled Process Stream Out

Reboiler Specifications

Unit:	E – 106, A/B
Type & Sizing	
Exchanger Type	Reboiler
Area	1313 ft^2
TEMA Type	AKU
Head Type	U Tube
Function	Vaporization of Liquid in Tanks to Send Back to Tower
Conditions	
Shell Fluid	DME/Water/Methanol
Tube Fluid	Low Pressure Steam
Max Operating Temperature	340 ^oF
Design Temperature	306 ^oF
Operating Pressure	172 psia
Design Pressure	222 psia
Nozzles/Connections	
Connection	Service
1	High Pressure Steam In
2	Condensate Stream Out
3	Cool Process Stream In
4	Hot Process Stream Out

Unit:	E – 202, A/B
Type & Sizing	
Exchanger Type	Reboiler
Area	12,280 ft^2
TEMA Type	AKU
Head Type	U Tube
Function	Vaporization of Liquid in Tanks to Send Back to Tower
Conditions	
Shell Fluid	Methanol/Water
Tube Fluid	Low Pressure Steam
Max Operating Temperature	380 ^oF
Design Temperature	345 ^oF
Operating Pressure	144 psia
Design Pressure	194 psia
Nozzles/Connections	
Connection	Service
1	High Pressure Steam In
2	Condensate Stream Out
3	Cool Process Stream In
4	Hot Process Stream Out

Tk – 101 A/B Specification Sheet

Unit:	TK-100 A/B
Type & Sizing	
Material	Austenitic 304L Stainless Steel
Shape	Cylindrical
Roof	Internal floating roof
Diameter	138 ft
Height	44 ft
Wall thickness	1.5 in
Conditions	
Chemical Storage	Methanol
Liquid Volume	4.89E+06 gal
Operating Temperature	77°F
Design Temperature	177°F
Operating Pressure	14.2 psia
Design Pressure	29.4 psia
Specific Gravity	0.787
Nozzles/Connections	
Connection	Service
1	Railcar offload inlet
2	Process feed stream outlet
3	Liquid overflow outlet

M – 301 Specification Sheet

Unit:	M-301
Type & Sizing	
Material	Carbon Steel
Orientation	Vertical
Diameter	4.7 ft
Length	14.1 ft
L/D	3
holdup time	10 minutes
Conditions	
Fluid in	DME, lubricant additive
Fluid out	DME with 900ppm by mass of lubricant
Operating Temperature	77°F
Design Temperature	177°F
Operating Pressure	75 psig
Design Pressure	125 psig
Volume flow	11,000 gal/hr
Power input	2.1 kW
Nozzles/Connections	
Connection	Service
1	DME inlet
2	Lubricant inlet
3	DME mixed outlet

Tk – 302 Specification Sheet

Unit:	TK-302
Type & Sizing	
Material	Carbon Steel
Orientation	Horizontal
Diameter	6.4 ft
Length	19.2 ft
Wall thickness	0.25 in
Conditions	
Chemical Storage	Lubricant addititve
Liquid Volume	4,603 gal
Operating Temperature	77°F
Design Temperature	177°F
Operating Pressure	0 psig
Design Pressure	50 psig
Specific Gravity	0.898
Nozzles/Connections	
Connection	Service
1	Lubricant inlet
2	Lubricant outlet to mixer

Equipment Cost Summary

Reactors

The reactor costs were calculated based on the cost correlations found in Turton for a vertical pressurized vessel (21). Costs are based on the volume of the vessel, the material of the vessel, and the design pressure for the vessel. All costs are based on 2001 dollars, and these are brought to mid-2017 using CEPCI values. Basic costs for the vessel itself utilize Equation 30, while pressure factors are calculated using Equation 31 and the total bare module factor is calculated with Equation 32. Finally, the present value of the equipment is calculated and designated with C_{TM} using Equation 34.

$$C_p^o = 10^{(K_1+K_2 Log(A)+K_3(Log(A)^2))} \quad 30$$

$$F_p = \frac{\frac{(P+1)D}{2(850-.6(P+1))}+0.00315}{0.0063} \quad 31$$

$$F_{BM} = (B_1 + B_2 F_p F_m) \quad 32$$

$$C_{BM} = F_{BM} C_p^o \quad 33$$

$$C_{TM,2017} = C_{BM} * \frac{CEPCI\ 2017}{CEPCI\ 2001} * Fees * Contingency \quad 34$$

Table 22: Reactor Initial Fixed Capital Costing Parameters

Variable	Value	Units	Calculated Variable	Value	Units
P	38.5	barg	A	3.21	m3
D	0.9144	m	Cpo	$ 5,650	2001 $
K1	3.4974	n/a	Fp	3.92	n/a
K2	0.4485	n/a	Fbm	9.388	n/a
K3	0.1074	n/a	Fm	1	n/a
B1	2.25	n/a	Reactor Cbm (2017)	$ 610,000	2017 $
B2	1.82	n/a	Ctm (2017)	$ 1,060,000	2017 $
CEPCI 2017	566.6	n/a	Fees	3%	n/a
CEPCI 2001	397	n/a	Contingency	15%	n/a

The catalyst was assumed to require replacement once annually, so the entire reactor volume of catalyst expense was considered as an annual operating cost. For 1.81 cubic meters of catalyst, the cost at $5.00 per pound was determined to be $17,600 per reactor. Utilizing the installation factors for the pressure vessels, the annual cost of catalyst replacement for all four reactors was determined to be $286,000. Assuming a 20-yeat plant life and a ROR of 8%, the total present worth cost (PWC) of all four reactors was $3.87 million. Table 23 displays the annual cost summary for the reactors.

Table 23: Annual Cost Summary Sheet for Reactors

Variable	Value	Units
Volume of Catalyst	1.81	m3
Cost of Catalyst	$5	$/lb
Replacement Schedule	12	months
Fbm	4.07	n/a
Annual Expense per Reactor	$ 286,000	$/year
PWC @8%ROR for 20 years	$ 2,810,000	2017 $

Though some reactor cost correlations were known, they were not utilized for a few reasons. While the provided cost correlations for a jacketed reactor produced a total module cost (Ctm) of $250,000, that does not explicitly account for the operating pressure. Furthermore, it does not account for the cost of the pipes inside the reactor, suggesting that it would be an underestimate. The pressure vessel correlations, however, produced a single reactor total module cost of $264,000, which accounts explicitly for the pressure design. As such, the total initial capital expense for the reactors was $1.06 million, and the PWC for the reactors and catalyst at 8% ROR over 20 years was $3.87 million.

Heat Exchangers

The heat exchangers were all costed according to the correlations in Turton. Costs are based on the heat transfer area of the exchanger, its material, and its design pressure. All costs are based on 2001 dollars, and these are brought to mid-2017 using CEPCI values. The cost equations utilized included Equations 30-34, while the pressure factor was calculated using Equation 35. A summary of costing values is shown in Table 24.

$$F_p = 10^{(C_1 + C_2 \mathrm{Log}(P) + C_3(\mathrm{Log}(P)^2))} \tag{35}$$

Table 24: HEX Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	40	Fm	1
Fp	1.1702	Fbm	3.572
K1 (Fixed Tube)	4.3247	C1	0.03881
K2 (Fixed Tube)	-0.303	C2	-0.11272
K3 (Fixed Tube)	0.1634	C3	0.08183
B1	1.63	Fees	3%
B2	1.66	Contingency	15%

Table 25: HEX Costing Results

Parameter	E-103 A/B	E-102 A/B	E-101 A/B	E-104 A/B	E-203 A/B
A (ft2/exchanger)	402	552	502	778	118
Total Area	804	1104	1004	1556	236
Total Cpo	$36,000	$39,000	$38,000	$21,000	$15,000
Total Ctm	$215,000	$232,000	$226,000	$255,000	$185,000
TEMA Type	AEL	AEL	AEL	AEL	AEL
Annual Expense				$220,000.00	$24,500.00
PWC @8% for 20 years	$215,000	$232,000	$226,000	$2,416,000	$426,000

The total capital expense for the heat exchangers was $1.11 million. Only HEX-4 A/B requires the use of cooling water. As such, the annual expenditures associated with E-103 A/B, 2, & 3 is purely maintenance. The annual cooling water expense for E-104 is calculated from the required flow rate of water. The cost of cooling water on an annual basis was calculated by multiplying the flow rate by the operating time per year (assuming a 0.95 service factor) and the average cost of cooling water in the Gulf Coast, which was assumed to be $0.05/m^3$. The annual cost of cooling water was determined to be $244,500, which corresponds to a total HEX PWC of $3.52 million dollars over 20 years at 8% ROR.

Furnace

The cost of the preheating furnace was estimated using the Guthrie cost correlations provided in Turton (21). The base cost was calculated with Equation 30 using the power supplied in kW. The pressure factor was calculated using Equation 35, while an additional bare module factor was calculated using Figure A.19 in Turton, which provides material bare module factors for furnaces and other equipment (21). Guthrie's correlation also requires a unique bare module calculation, which is provided as Equation 36. An additional 3% was included for fees and 15% for contingency, producing a 2017 total module cost of $3.30 million. Table 25 displays the costing results.

Furnace Bare Module Cost $$C_{BM} = C_p^o F_p F_T \qquad (36)$$

Table 26: Furnace Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	40	Ft	1
Fp	1.097	Fbm	2.1
K1	7.3488	C1	0.1405
K2	-1.1666	C2	-0.2698
K3	0.2028	C3	0.1293
Contingency	15%	Fees	3%

The annual operating cost of the furnace was calculated based on the fuel gas flow rate requirements. As the furnace operates at a duty of 10.6 MW, assuming a stream factor of 0.95, the annual energy required to preheat the stream is 3.04×10^{11} kJ per year, which equates to 3.40×10^{8} cubic feet of fuel per year. At a cost of $3.50 per 1,000 cubic feet, the annual cost of fuel is $1.19 million dollars per year. Table 26 provides a summary of furnace costing.

Table 27: Furnace Costing Results

Parameter	Value
Q (BTU/hr)	3.48E7
Cpo	$ 2,800,000
Ctm	$ 3,300,000
Annual Expense	$ 1,190,000
PWC @8% ROR for 20 years	$ 11,700,000

The furnace is one of the most expensive pieces of equipment in the process, and the annual cost of fuel is a significant expense. While eliminating it would be beneficial to cost reduction, the energy required to vaporize the methanol feed is so significant that a furnace is the only practical way to achieve it at the design conditions. As such, though it has a significant impact on the overall cost of the process, its function is required and has been retained in the design.

Towers

Towers T-101 A/B and T-201 A/B were priced using cost correlations from Turton. Key parameters that affected costing were design pressure, volume of the tower, and material of construction. All costs are based on 2001 dollars, and these are brought to mid-2017 using CEPCI values. The cost equations utilized are Equations 30-34. A summary of costing values for both towers are shown in Table 28 and Table 29.

Table 28: T-101 Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	14.3	Fm	1
Fp	1.82	Fbm	5.56
K1 (Fixed Tube)	3.4974	C1	N/A
K2 (Fixed Tube)	0.4485	C2	N/A
K3 (Fixed Tube)	0.1074	C3	N/A
B1	2.25	Fees	3%
B2	1.82	Contingency	15%

Table 29: T-201 Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	12.35	Fm	1
Fp	1.65	Fbm	5.25
K1 (Fixed Tube)	3.4974	C1	N/A
K2 (Fixed Tube)	0.4485	C2	N/A
K3 (Fixed Tube)	0.1074	C3	N/A
B1	2.25	Fees	3%
B2	1.82	Contingency	15%

The Costing Results for T-101 A/B and T-201 A/B are shown in Table 30 below. The total capital expense for the towers is $552,000. Annual expenditures towards the towers are only due to maintenance.

Table 30: Tower Costing Results

Parameter	T-101 A/B	T-201 A/B
Volume (ft^3/per tower)	452.0	630
Total Area for A/B	904.0	1260
Total Cpo	$26,800	$34,000
Total Ctm 2017	$252,000	$300,000

Trays

The trays for T-101 A/B and T-202 A/B were priced using cost correlations from Turton. The area of the trays, tray type, and number of trays were the key parameters for determining the price of the trays. Equation 30 was used to calculate Cpo for an individual tray and then Cpo was multiplied by the number of trays to find the Total Cpo. Equation 34 was used to escalate Ctm from 2001 dollars to mid 2017 dollars using CEPCI values for each year. A summary of costing values are shown in Table 31 and Table 32.

Table 31: T-101 Tray Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	N/A	Fm	N/A
Fp	N/A	Fbm	1
K1 (Fixed Tube)	2.9949	C1	N/A
K2 (Fixed Tube)	0.4465	C2	N/A
K3 (Fixed Tube)	0.3961	C3	N/A
B1	N/A	Fees	3%
B2	N/A	Contingency	15%

Table 32: T-201 Tray Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	N/A	Fm	N/A
Fp	N/A	Fbm	1
K1 (Fixed Tube)	2.9949	C1	N/A
K2 (Fixed Tube)	0.4465	C2	N/A
K3 (Fixed Tube)	0.3961	C3	N/A
B1	N/A	Fees	3%
B2	N/A	Contingency	15%

The Costing results for the trays of T-101 A/B and T-201 A/B are displayed in Table 32 below. The total capital expense for the trays is $207,000. The only annual expenditure for the trays is maintenance.

Table 33: T-101 Tray Costing Results

Parameter	Trays for T-101 A/B	Trays for T-201 A/B
A (ft2/per tray)	7.1	7.1
Total Area for A/B	14.1	14.1
Number of trays	27	45
Total Cpo	$1,700	$1,700
Total Ctm 2017	$77,000	$130,000
Tray Type	Sieve Tray	Sieve Tray

Condenser and Reboilers

The condesers, E-105 A/B and E-201 A/B, and the reboilers, E-106 A/B and E-202, were costed using the same heat transfer techniques as described in the Heat Exchangers section. A summary of costing values is shown in Table 34, Table 35, Table 36, Table 37, and Table 38.

Table 34:E-105 Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	14.1	Fm	1
Fp	1.041	Fbm	6.29
K1 (Fixed Tube)	4.3247	C1	0.03881
K2 (Fixed Tube)	-0.303	C2	-0.11272
K3 (Fixed Tube)	0.1643	C3	0.08183
B1	1.63	Fees	3%
B2	1.66	Contingency	15%

Table 35: E-106 Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	14.3	Fm	1
Fp	1.04	Fbm	6.3
K1 (Fixed Tube)	4.1884	C1	0.03881
K2 (Fixed Tube)	-0.2503	C2	-0.11272
K3 (Fixed Tube)	0.1974	C3	0.08183
B1	1.63	Fees	3%
B2	1.66	Contingency	15%

Table 36: E-201 Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	12.1	Fm	1
Fp	1.041	Fbm	6.29
K1 (Fixed Tube)	4.3247	C1	0.03881
K2 (Fixed Tube)	-0.303	C2	-0.11272
K3 (Fixed Tube)	0.1643	C3	0.08183
B1	1.63	Fees	3%
B2	1.66	Contingency	15%

Table 37: E-202 Costing Parameters

Parameter	Value	Parameter	Value
Design Pressure (barg)	12.35	Fm	1
Fp	1.04	Fbm	6.3
K1 (Fixed Tube)	4.1884	C1	0.03881
K2 (Fixed Tube)	-0.2503	C2	-0.11272
K3 (Fixed Tube)	0.1974	C3	0.08183
B1	1.63	Fees	3%
B2	1.66	Contingency	15%

Table 38: Reboiler and Condenser Costs

Parameter	E-105 A/B	E-106 A/B	E-201 A/B	E-202 A/B
A (ft2/per exchanger)	2131	1313	567	12280
Total Area for A/B	4262	2626	1134	24560
Total Cpo	$62,000	$68,000	$39,000	$370,000
Total Ctm 2017	$666,000	$716,000	$410,000	$3,920,000
TEMA Type	AEL	AKU	AEL	AKU
Total Annual Expense	$188,000	$476,000	$350,000	$3,440,000
PWC @8% for 20 years	$2,500,000	$5,400,000	$3,900,000	$38,000,000

The total capital expense for the condensers, E-105 and E-201, are $1,076,000. The annual expense amounts to $538,000, which is the cost of cooling water per year. The total capital expense for the Reboilers, E-106 and E-202, is $4,640,000. The annual expenses for the reboilers amount to $3,920,000 which is the price of low pressure steam. Steam was priced at $30 dollars per 1000 kg as suggested by Turton. This leads to an overall PWC, for the reboilers and the condensers, of $49,800,000 over 20 years at an 8% ROR.

Pumps

Every pump was priced based on cost correlations from Turton (21) and the electricity rates were based on the electricity rates in the Lake Charles, Louisiana area. Pricing of pumps depends on the design pressure, the brake horse power, and the hydraulic power. The Turton correlation costs the pumps in 2001 dollars and they are brought to mid-2017 dollars using the CEPCI index. Equations 30 and 32-35 were used for costing. The Material of Construction factor for carbon steel was found to be 1.6 for centrifugal pumps and 1.4 for positive displacement pumps based on Figure A.18 in Turton (21).

Table 39: Parameters for Centrifugal Pumps in System

Parameter	Value	Parameter	Value
Fm	1.6	C1	-0.3935
K1	3.3892	C2	0.3957
K2	0.0536	C3	-0.00226
K3	0.1538	Fees	3%
B1	1.89	Contingency	15%
B2	1.35		

Table 40: Parameters for Positive Displacement Pump in System

Parameter	Value	Parameter	Value
Fm	1.4	C1	0
K1	3.4771	C2	0
K2	0.135	C3	0
K3	0.1438	Fees	3%
B1	1.89	Contingency	15%
B2	1.35		

Table 41: Pump Costing Results

Parameter	P – 101, A/B/C/D	P – 102, A/B/C/D	P – 103, A/B/C/D	P – 201, A/B/C/D	P – 301, A/B
Hydraulic Power (kW)	33.2	1.11	0.29	1.16	0.0031
Total Hydraulic Power (kW)	66.4	2.22	0.58	2.32	0.0062
Brake Horse Power (kW)	66.4	2.22	0.82	2.32	0.0089
Total Brake Horse Power (kW)	132.8	4.44	1.64	4.64	0.0178
Fp	1.67	1.14	1.13	1.10	1
Total Cpo	$ 12,000	$ 10,000	$ 10,000	$ 10,000	$ 6,000
Total Ctm	$ 114,000	$ 75,000	$ 71,000	$ 74,000	$ 20,000
Annual Expense	$ 49,000	$ 2,000	$ 1,000	$ 2,000	$ Negligible
PWC @8% for 20 years	$ 592,000	$ 91,000	$ 77,000	$ 90,000	$ 20,000

The total cost for the pumps in the process is $354,000. The annual expense for operating the pumps includes electricity costs and any maintenance. The annual expense of P – 301 is assumed to negligible because of its low costs compared to the other pumps in the system. The electricity costs were calculated multiplying the the brake horsepower of the pumps by the amount of hours in a year, while also including a 0.95 service factor. The average cost of industrial electricity in the Lake Charles area is $0.0441/kWhr (29). The total annual costs of the pumps were calculated to be $54,000, this correlates to a total PWC of $870,000 over 20 years at 8% ROR.

Condensate Receivers

The condensate receivers were priced using the parameters and correlations from Turton. Pricing is based on the design pressure, the volume of the vessel, and the material of construction. Turton correlations costs in 2001 dollars and then converted to mid-2017 dollars using the CEPCI index. Equations 30-34 were used for the costing correlations and pressure factors. Table 42 and Table 43 shows all of the parameters required to cost the receivers.

Table 42: Condensate Receiver Parameters

Parameter	Value	Parameter	Value
Fm	1	B1	1.49
K1	3.5565	B2	1.52
K2	0.3776	Fees	3%
K3	0.0905	Contingency	15%

Table 43: Condensate Receiver Results

Parameter	V – 101, A/B	V – 201, A/B
Volume (ft^3)	200.2	127.9
Diameter (ft)	4.40	3.79
Length (ft)	13.19	3.46
Pressure (psia)	203.7	174.7
Fp	1.77	1.31
Total Cpo	$ 16,000	$ 13,000
Total Ctm	$ 77,000	$ 51,000
Annual Expense	$ N/A	$ N/A
PWC @8% for 20 years	$ 77,000	$ 51,000

A total module cost of $128,000 was calculated for the four total condensate receivers. There are no annual expenses for the condensate receivers other than maintenance. Since there are no annual expenses with the condensate receivers, the total PWC is $128,000 for 20 years at 8% ROR.

Storage Tanks

Using the Turton cost correlations, the cost of each storage tank was estimated. The primary design parameters in these cost estimates are dependent on the total design volume, the operating pressure, and the material of construction. The base cost, C_p^o was calculated using Equation 30 and the total volume in cubic meters being the capacity factor. Once the base cost was found, the pressure factor, F_p, and material factor, F_m, were found. Though the tank operates at atmospheric pressure, a pressure factor of 1.25 was included to account for its large size. A material factor was found to be 3.1 for a stainless steel tank. The costing parameters as well as the correlation constants can be found below in Table 43. Using Equation 33, the bare module cost, C_{BM}, was estimated. In order to accurately estimate the total cost of the storage tanks, the bare module cost of the tanks was multiplied by an index ratio in order to bring the cost to terms of late-2017 dollars. From this updated cost, an 18% increase due to contingency and fees was added to the bare module cost in order to estimate the total module cost, C_{TM}.

Table 44: Costing parameters and values for methanol feed storage tanks

Parameter	Value
Volume based on dimensions (m^3/tank)	**18,636**
K1	**5.9567**
K2	**-0.7585**
K3	**0.1749**
C_p^0 ($)	**807,418**
P (barg)	**-0.034**
Fp	**1.25**
Fm	**3.1**
B1	**2.25**
B2	**1.82**
Cbm/tank ($/tank)	**7,511,008**
Cbm both tanks ($)	**15,022,015**
Cbm both tanks (2017 $)	**22,022,198**
Ctm both tanks(2017 $)	**25,986,194**
Ctm/tank (2017 $/tank)	**12,993,097**

Mixer

The mixer was costed as a sum of the impeller and motor as well as for the mixing vessel itself. First the impeller base cost was determined using the power required and Equation 30. The bare module factor, F_{BM}, was assumed to be 1.38. The bare module cost was then determined by multiplying the base cost with the bare module factor. From the bare module cost, a combined 18% was added to account for contingency and fees resulting in the total module cost, C_{tm}. The C_{tm} was then brought to 2017 dollars through a ratio of costing indices to give a better estimate of the actual cost. The mixer impeller and motor cost values can be found in Table 44 below.

Table 45: Costing parameters and values for mixer impeller motor

Parameter	Value
Mixer tank volume (m^3)	6.92
Impeller motor power input (kW)	2.0760
K1	3.8511
K2	0.7009
K3	-0.0003
C_p^o	11,842
F_{bm}	1.38
C_{bm}	16,342
C_{tm}	19,357
2017 C_{tm}	27,626

The tank was costed using the same equations. For the mixing vessel, the capacity factor used in determining the base cost was the design volume. A bare module factor of 4 was given in the Turton Table A.7 for a mixer. Using the bare module factor and base cost, the bare module cost was found by multiplying the two factors together. Similarly, to the impeller and motor cost, an 18% increase due to contingency and fees was added to the bare module cost in order to find the total module cost. This C_{tm}, was also brought to 2017 dollars to better estimate the total cost of the mixer tank. Values used for the mixer tank costing correlations can be seen in Table 45.

Table 46: Costing parameters and values for mixer tank

Parameter	Value
Volume (m^3)	6.92
K1	4.7116
K2	0.4479
K3	0.0004
C_p^o ($)	122,508
F_{bm}	4
C_{bm} ($)	490,034
C_{tm} ($)	580,445
2017 C_{tm} ($)	828,413

The final costs associated with the mixing tank are the operating costs of the motor. Assuming a 0.95 service factor, the annual hours of operation for the motor were found to be 8,322 hr/yr. The electricity costs for industrial use in Lake Charles, LA was found to be 0.0441 $/(kWh). Using this price for electricity and the annual power consumption of the motor, the annual electricity cost was found to be $762/yr.

Table 47: Costing parameters and values for annual electricity costs of mixer motor

Parameter	Value
Annual hours of use	**8,322**
Cost of Electricity $/kWh	**0.0441**
Annual Cost ($)	**762**

Lubricant Storage Tank

The lubricant tank cost was estimated using costing correlations found in Turton Appendix A. The base cost, C_p^o, was found using Equation 30 with the design volume as the capacity factor. An F_p factor of 1 was used since the minimum thickness of the tank is larger than the thickness required for the given pressure and diameter. Using carbon steel as the material, a material factor, F_m, of 1. Using the constants found in Table 47, the bare module cost for the lubricant tank was found to be $46,700. A 15% contingency factor and a 3% fees factor were then added onto the bare module cost to find the total module cost, C_{tm}. With the total cost being found, the total cost value was then brought forward to 2017 dollars bring the actual cost of the tank to approximately $79,000. All constants and values used in the tank costing estimates can be seen in Table 47.

Table 48: Costing parameters and values for lubricant storage tanks

Parameter	Value
Volume based on dimmensions (m^3/tank)	**19.2**
K1	**3.5565**
K2	**0.3776**
K3	**0.0905**
C_p^0 ($)	**15,476**
P (barg)	**3.6**
Fp	**1**
Fbm	**3.01**
Cbm ($)	**46,582**
Ctm ($)	**55,176**
2017 Ctm ($)	**78,748**

Fixed Capital Investment Summary

Table 49: Summary of Equipment Investments

Rounded Totals		58	$ 30,000,000	$ 54,300,000		$ 6,230,000	$ 115,500,000
Equipment	**Ctm per unit**	**Units**	**Cbm Total**	**Ctm Total**	**Cost/unit/yr**	**Cost/yr**	**20% PWC @ 8%**
R-101 A/B/C/D	$ 260,000	4	$ 890,000	$ 1,060,000	$ 70,000	$ 290,000	$ 3,870,000
H-101	$ 3,300,000	1	$ 2,790,000	$ 3,300,000	$ 1,190,000	$ 1,190,000	$ 14,980,000
E-101 A/B	$ 110,000	2	$ 190,000	$ 230,000	$ -	$ -	$ 230,000
E-102 A/B	$ 120,000	2	$ 200,000	$ 230,000	$ -	$ -	$ 230,000
E-103 A/B	$ 110,000	2	$ 180,000	$ 220,000	$ -	$ -	$ 220,000
E-104 A/B	$ 130,000	2	$ 220,000	$ 260,000	$ 110,000	$ 220,000	$ 2,420,000
E-105 A/B	$ 330,000	2	$ 560,000	$ 670,000	$ 90,000	$ 190,000	$ 2,520,000
E-106 A/B	$ 360,000	2	$ 600,000	$ 720,000	$ 240,000	$ 480,000	$ 5,400,000
T-101 A/B Shell	$ 130,000	2	$ 210,000	$ 250,000	$ -	$ -	$ 250,000
T-101 A/B Trays	$ 40,000	2	$ 70,000	$ 80,000	$ -	$ -	$ 80,000
V-101 A/B	$ 40,000	2	$ 70,000	$ 80,000	$ -	$ -	$ 80,000
E-201 A/B	$ 210,000	2	$ 350,000	$ 410,000	$ 180,000	$ 350,000	$ 3,870,000
E-202 A/B	$ 1,960,000	2	$ 3,310,000	$ 3,920,000	$ 1,720,000	$ 3,440,000	$ 37,700,000
E-203 A/B	$ 90,000	2	$ 160,000	$ 180,000	$ 10,000	$ 20,000	$ 430,000
T-201 A/B Shell	$ 150,000	2	$ 250,000	$ 300,000	$ -	$ -	$ 300,000
T-201 A/B Trays	$ 60,000	2	$ 110,000	$ 130,000	$ -	$ -	$ 130,000
V-201 A/B	$ 30,000	2	$ 40,000	$ 50,000	$ -	$ -	$ 50,000
M-301	$ 830,000	1	$ 700,000	$ 830,000	$ -	$ -	$ 840,000
Tk-302	$ 80,000	1	$ 70,000	$ 80,000	$ -	$ -	$ 80,000
P-101 A/B/C/D	$ 30,000	4	$ 100,000	$ 110,000	$ 20,000	$ 50,000	$ 590,000
P-102 A/B/C/D	$ 20,000	4	$ 60,000	$ 80,000	$ -	$ -	$ 90,000
P-103 A/B/C/D	$ 20,000	4	$ 60,000	$ 70,000	$ -	$ -	$ 80,000
P-201 A/B/C/D	$ 20,000	4	$ 60,000	$ 70,000	$ -	$ -	$ 90,000
P-301 A/B	$ 10,000	2	$ 20,000	$ 20,000	$ -	$ -	$ 20,000
Tk-100 A/B	$ 13,000,000	2	$ 18,700,000	$ 26,000,000	$ -	$ -	$ 26,000,000
Grassroots Cost	$ 15,000,000	1	$ -	$ 15,000,000	$ -	$ -	$ 15,000,000

Safety, Healthy, and Environmental Considerations

The proposed design has been analyzed for hazards and a number of safety steps have been taken to reduce the likelihood of an accident occurring. In addition to routine maintenance, proper process safety management (PSM), safety culture, safe startup and shutdown procedures, and development of a RMP, special considerations for the following has been and must be continued to be considered as the project advances.

Overall Site Safety

The location of the plant is within a reasonably well populated portion of Lake Charles, LA. As such, there is the potential to cause widespread damage in case of a toxic release or an explosion. Despite these hazards, the location is good for access to sewers, trucks, and tracks for railcar feed. As the project advances there should be studies done on the effects of releases and explosions in the area, and required greenspace must be purchased as an exclusionary zone outside the plant.

Storage Tanks

When working with a highly reactive chemical such as methanol multiple safety and environmental hazards need to be considered. The storage tanks have been designed such that the thickness of the tanks can withstand the internal pressures exerted by the volume of liquid being stored. In case a fire, an alcohol resistant firefighting system will be installed in order to properly extinguish a methanol fire. A general guideline for potential loss of containment of a hydrocarbon (21) states that a secondary spill area should equal 110% of the working volume of the tank. Since each tank will have a working volume of up to 4.10 x 10^6 gallons, a very large containment area will need to be in place. To account for this spill capacity, the tanks will sit 10ft into the ground with a rectangular spill area surrounding the tanks. This area will have the dimensions of 220ft x 275ft in order to properly hold the capacity of a complete storage tank spill scenario. Another factor that was considered when designing the tanks was the wind. Louisiana is an area commonly susceptible to hurricanes. To account for potential hurricane winds, the storage tanks were designed with a much larger diameter than height to prevent a scenario where the tanks for suffer from a top-heavy rupture in the walls.

Pumps

The pumps were designed to be inherently safe for this process. Every pump was designed to handle 50 psi above its operating pressure. When calculating the hydraulic power of the pumps, a 10 % safety factor was used to ensure the pump could handle any surges that occur throughout the system. The metering pump allows for a precise amount of lubricant that will be added to the mixer. Not having the proper amount of lubricant in the DME fuel can have consequences and lead to engine damage. The pumps should be regularly checked for preventive maintenance and to ensure the pumps are operating properly. Spares are also installed in parallel to the pumps in operation. Spares allow the system to continually operate so the system does not need to shut down if any pump is broken.

Heat Exchangers

Heat exchangers are inherently safe to use and operate. They are designed to handle temperatures and pressures well above what should be seen in normal operation of the process. They have similar sizes for outlet and inlet connections, thus reducing the possibility of an overpressure occurring within the exchanger itself. The heat exchangers in the proposed process operate at high temperatures, so they must all be properly insulated to prevent injury and to reduce ambient heat loss to the atmosphere. In addition, the heat exchangers should regularly be examined for corrosion and fouling.

Reactors

The reactors utilized in this process are pressurized, adiabatic vessels that contain a highly exothermic chemical reaction. As such, they have a number of potential hazards that they pose to the process and the operators. The reactors operate at an outlet temperature of 720°F and a pressure of 500 psia, which have the potential to cause harm if containment is lost. In addition, the reaction is an exothermic reaction that has the potential to runaway and create a massive amount of pressure within the vessel, potentially causing an explosion and release of methanol and DME.

A number of safety measures are proposed to mitigate the risk associated with these hazards. The most significant safety factor built in to the design was splitting the process into four smaller reactors rather than a single larger reactor. This minimizes the size of the reactors and the potential consequences of a undesired release. It also allows routine maintenance to be conducted on a reactor while the plant operates at 75% capacity, decreasing the capacity that the plant loses when a process unit goes down.

Proper jacketing on the reactor can reduce the probability that contact with the tank will burn an operator and will ensure the reactor maintains the energy released to increase the rate of reaction. A pressure relief valve will be installed on the vessel to prevent over pressuring the tank and causing an explosion. The discharge from the pressure relief will be sent to a flare to prevent accidental discharge of DME and MeOH to the atmosphere even in rare conditions.

Heat Exchangers

The possibility of a runaway reaction is unlikely, but it poses a potentially dangerous consequence if it does occur. In the case that it does occur, the final heat exchanger has been equipped with a bypass, which will drastically reduce the inlet temperature of the vapor and greatly slow the reaction to a controllable rate. If this does not provide control of the system, the furnace can be bypassed as well, and if necessary the entire system can be shut down at the inlet feed pumps. With the reactants cut off, the reaction will soon die out and damage to the system or loss of containment will be avoided. Though the heat exchangers are contain a number of hazards, control can be maintained and safety can be prioritized through provided process controls.

Furnace

The furnace also contains a number of potential hazards, including, but not limited, to high temperature and pressure, flammability, and explosivity. Temperature hazards can be mitigated through measures described above for the heat exchangers and reactors, including both insulation and proper materials of construction. To avoid damage to the furnace and reduce the amount of startup/shutdown time experienced, a stream has been included to bypass the furnace to take it offline without requiring total furnace shutdown. The temperature of the furnace is controlled through adjusting the flow rate of the inlet fuel gas, so in case of an increase in temperature and pressure within the process stream, the temperature of the furnace can be reduced.

Towers

Routine inspection and maintenance is needed for the distillation columns as this prevents against corrosion and potential leaks. It is important to have a pressure relief system in place in the case that the tower becomes over pressurized to prevent a possible explosion. A loss of cooling water to E-104, E-105, or E-201 could pose a hazard. If cooling water is lost in E-104, the mixture entering T-101 will be at the incorrect temperature, which could result in increased flowrates in order to maintain pressure inside the column, resulting in incorrect product compositions. If E-105 or E-201 lose cooling water, the steam exiting T-101/T-201 will remain in the vapor phase, causing an over pressurization of the system potentially which could lead to malfunction/burst in the system (30).

Vessels

Every condensate receiver was designed to be safe to the environment and for the workers in the plant. The vessels were sized to hold 50 psi above their operating pressure. This ensures that if any pressure fluctuations occur that the tank will be safely able to handle it. Also, the upper limit of the suggested hold up time of 10 minutes was used for the volume calculations.

Mixer

With the size of the mixer and the purpose it is providing for the process, there are not many safety and environmental concerns that need special attention. The mixer is made out of carbon steel which will need an annual inspection to check for any potential corrosion in the tank. Also, the impeller was designed with the viscosities of the lubricant additives and DME in mind in order to eliminate any possible failures associated with mixing too vigorously.

Wastewater

The wastewater from the process was specified so as to keep it below the lower flammability limit (LFL) of methanol in the vapor coming off of the water. The LFL for methanol is 6.7% by volume, so the water composition of the waste stream was specified to be 0.995 by mole composition, which produced a methanol vapor volume fraction of 0.0544, which is below the LFL and leaves some room for operational cushion without over specifying the product. The wastewater is supplied to a local gravity sewer, which would dilute the methanol concentration further and eliminate any hazard of flammability or explosion while it is sent through the collection system to the wastewater treatment plant.

Other Important Considerations

Environmental Considerations

A number of environmental concerns may be raised when considering the implementation of the proposed design. As is the case with most chemical industries, transportation provides an increased possibility for accidental discharge of a chemical into the environment. The proposed site is located within a few miles of the methanol production plant that the feedstock will be acquired from, reducing the time that the chemical is in transit and also the chance of an accident occurring en route to the facility.

In addition, the use of the furnace to preheat the feed stream and the production of high-pressure steam for the reboilers of the distillation columns requires the burning of fossil fuels, which is known to release carbon dioxide into the atmosphere. This and other unintended vapor releases can contribute to air pollution, which can have a negative impact on the air quality in the region.

Finally, in the unlikely event that any loss of containment would take place, the location of the plant adjacent to a body of water that connects to a network of surface waterways suggests that the transmission of any chemical leaked into the water could quickly spread to the surrounding waters. This poses a hazard for the wildlife in the area as well as the habitat and the humans living around the plant. The likelihood of an accidental discharge that impacts the connecting waterways is highly unlikely, so environmental factors do not eliminate the current lot from consideration.

Methanol is harmful to the environment and all releases should be avoided. DME has fewer negative environmental impacts than diesel, but containment is always a priority. Gas detection sensors and sniffers will be used to detect any gasses being released to the environment. Methanol in wastewater is an environmental hazard if not removed to a safe level. The methanol was removed to be under the Lower Explosive Limit and is much safer to the environment. Berms will be placed around any storage vessel in case of accidental releases of chemicals to ensure containment and prevent release into the environment.

Despite these environmental hazards, DME is a clean alternative to diesel fuel. Therefore, the environmental hazards associated with the facility are minimal. A DME spill will be much less hazardous to the environment compared to diesel. Less carbon emissions will be released into the environment when diesel engines are replaced with DME engines.

Health and Safety Consideration

Safety

DME and methanol are extremely flammable. Extreme caution will be put in place to remove any ignition sources where DME and methanol vapors may be present. DME and methanol are heavier than air and can settle in locations with low ventilation. Local exhaust ventilation will be installed to prevent any settling of hazardous vapors and for a safer air for workers to breathe. General exhaust systems will be used where heat control would be beneficial. Including fire suppression systems within the facility would help prevent and extinguish any fires that could occur.

The high volume of trucks that are projected to be on or around the premises of the plant has the potential to provide sources of ignition for any methanol or DME vapor clouds. As such, the proposed filling station will be placed far enough away from the proposed process to ensure reduced risk of ignition in case of a loss of containment. Furthermore, high volumes of traffic can be hazardous to workers on foot in the area, so high visibility clothing and standard safety measures will be enacted at all times.

Health

There are several health considerations that must be taken into account for the DME production system. DME can cause frostbite if contacted with the liquid form, so direct contact with DME should be avoided. Direct inhalation of DME can cause blurred vision, headaches, and possible loss of consciousness. Methanol has similar health hazards to DME. Drowsiness, headache, and eye irritation are all symptoms of exposure to methanol vapors. High concentration exposure can cause blindness and or even death. Personal protective equipment will be worn in areas where DME and methanol vapor may be present to reduce the risk of exposure.

HAZOP

Table 50: HAZOP Study Findings

No.	Guide Word	Deviation	Causes	Consequences	Safeguards	Recommendations
1	More	Pressure increase in T-101/201	Loss of CW to T-101/201	Column failure, rupture of lines, explosion	Pressure release valve on condensate receiver	Install a secondary relief device to reduce probability of discharge through main discharge valve
2	As well as	Two-phase flow to P-101	Energy from recycle and drop in pressure causes liquid to flash	Cavitation in P-101	Temperature and pressure controls on T-201	Investigate recycle stream connection methods
3	Reverse	Back flow to Tk-100	Flow from Recycle back into Tk-100 A/B	Increase in temperature and pressure in Tk-100	Check valve to prevent back-flow	Determine best method to utilize energy from recycle stream while preventing back flow
4	No	Liquid seal in E-106 and E-202 lost	Failure of control system to maintain level	Cavitation in P-103, loss of methanol to waste stream	High and low liquid level alarms on the control display	Consider use of panel alarms or secondary level measurement systems
5	More	Runaway reaction in R-101	Increase in feed temperature	Deactivation of catalyst, increase in pressure in R-101, explosion, release	Temperature alarms entering R-101, panel alarms if pressure or temperature gets too high	Ensure proper function of emergency bypass streams to prevent furnace overheating

Process Control Strategy

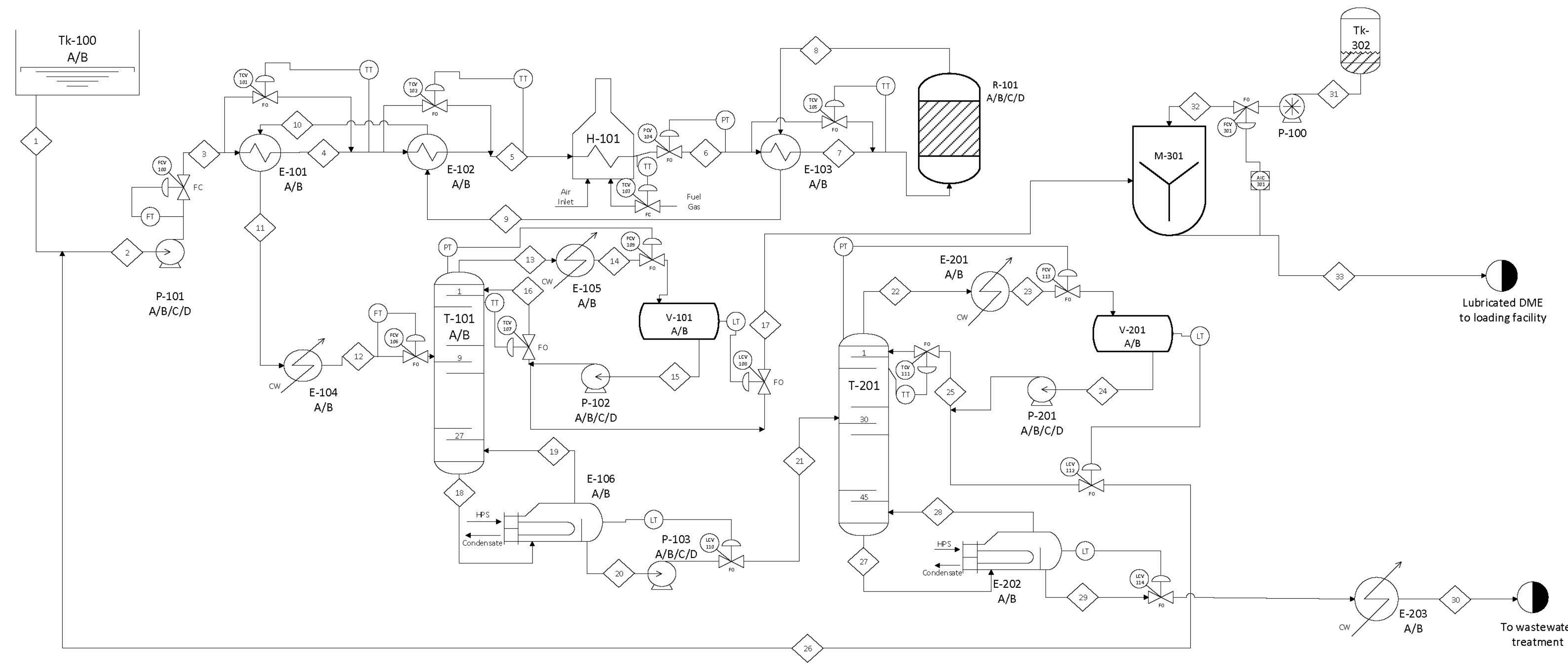
Table 51: Control Strategy, CV/MV Pairings, and Valve Failure Positions

Loop Tag	CV	MV	Valve Failure Position
FCV-100	Flow rate in Stream 3, P-1 to E-101 A/B	Flow rate in Stream 3, P-1 to E-101 A/B	Closed
TCV-101	Temperature of Stream 4, outlet of E-101 A/B	Flow rate of preheat stream bypassing E-101 A/B	Open
TCV-102	Temperature of Stream 5, outlet of E-102 A/B	Flow rate of preheat stream bypassing E-102 A/B	Open
TCV-103	Temperature in Stream 6, outlet of H-101	Fuel gas flow to H-101	Closed
PCV-104	Pressure in H-101 outlet, Stream 6	Stream 6 vapor flow rate, exiting H-101	Open
TCV-105	Temperature of Stream 7 at outlet of E-103 A/B	Flow rate of preheat stream bypassing E-103 A/B	Open
FCV-106	Flow rate of Stream 12 to T-101 A/B	Flow rate of Stream 12 to T-101 A/B	Open
TCV-107	Temperature in T-101 A/B	Stream 16 reflux flow rate to T-101 A/B	Open
LCV-108	Liquid level in V-101 A/B	Flow rate of Stream 17 from V-101 A/B to M-101	Open
PCV-109	Pressure in T-101 A/B	Flow rate of Stream 14 between E-105 A/B and V-101 A/B	Open
LCV-110	Liquid level in E-106 A/B	Flow rate of Stream 21 to T-201 A/B	Open
TCV-111	Temperature of T-201 A/B	Stream 25 reflux flow rate from V-201 A/B to T-201 A/B	Open
LCV-112	Liquid level in V-201 A/B	Recycle flow rate of Stream 26	Open
PCV-113	Pressure in T-201 A/B	Stream 23 flow rate between E-201 A/B and V-201 A/B	Open
LCV-114	Liquid level in E-202 A/B	Flow rate of wastewater in Stream 30	Open

The following page displays the general control strategy on a PFD. Not all piping and instrumentation is shown, as this is designed to give an idea of the basic strategy for controlling the process.

Tk-100 A/B	P-101 A/B/C/D	E-101 A/B	E-102 A/B	H-101	R-101 A/B/C/D	E-104 A/B	T-101 A/B	E-105 A/B	V-101 A/B	P-102 A/B	E-106 A/B	P-103 A/B	T-201 A/B	E-201 A/B	V-201	P-201 A/B	E-202 A/B
Parallel methanol storage tanks	Parallel Feed Pumps	Parallel feed Pre-heater	Parallel feed preheaters	Feed vaporizing furnace	Parallel methanol dehydration reactors	Parallel process coolers	Parallel DME separation tanks	Parallel DME condensers	Parallel DME reflux drum	Parallel reflux pumps	Parallel DME reboilers	Parallel bottoms/feed pumps	Parallel Meoh/ H2O separation columns	Parallel MeOH condensers	Parallel MeOH reflux drums	Parallel reflux pumps	Parallel H2O reboilers

P-202 A/B	M-301	P-301	Tk-302
Parallel wastewater pumps	DME and lubricant mixer	Lubricant pump to mixer	Lubricant storage tank



Startup

While the process should operate at steady state for the majority of the time, startup is a necessary transient condition that will affect the overall operation of the system. At the beginning of startup, T-101 and T-201 will be started in total reflux operation. For this to occur, a mixture of DME, methanol, and water is added to E-106 and steam and cooling water flow is started, but the feed and product streams remain closed. Likewise, E-202 will be seeded with a mixture of methanol and water. The steam and cooling water for the column will be started while the feed and outlet streams are shut.

During continuous operation, the majority of the heat transferred to the methanol reactor feed is done through heat integration processes in E-101, 102, and 103. In startup, however, the furnace will have to provide the heat required to elevate the inlet stream temperature to the 536°F required for optimal reactor use. For this to occur, the flow rate of fuel gas to H-101 will be increased to maximize the energy transferred to the stream while only one of the pumps P-101 will operate in order to reduce the overall flow rate to half of the design capacity.

The half flow stream will exit H-101 and enter R-101 A/B. This stream will be at lower pressure than it is during steady state operation, so conversion will be lower. This outlet stream will flow through the shell sides of E-101 ,102, 103, and 104. This stream will enter T-101, which has been operating at total reflux. Once the feed is added, distillate and bottom products will begin to be removed. With the removal of top and bottom products, DME will begin to flow to M-101 and the water/methanol mixture of Stream 20 will be pumped to T-201. The methanol and water exiting E-106 will be fed to T-201, which is operating at total reflux. As before, this will result in product streams exiting the unit, getting the process up to steady state.

With the heated product stream exiting R-101, the furnace will no longer be required to produce such high heat transfer, and the fuel gas rate can be reduced. Once the system is operating at half capacity through the parallel unit operations, the process can be started for the other side again starting with P-101 B.

The primary startup cost considerations are the increased fuel gas flow rate to H-101 required to preheat the process stream and the liquid mixtures for E-106 and E-202 that are required to start the columns operating at total reflux. These liquids will be obtained by saving them after shutdown on previous operations, in which case storage facilities will be needed, or purchased before initial startup, which requires additional working capital. Any costs associated with startup are assumed to be included in the costs for working capital, the total module cost, and contingency.

Manufacturing Costs (exclusive of Capital Requirements)

The total cost of operation was calculated using the method laid out in Turton Chapter 8. Raw materials were calculated based on purchase and transportation price of methanol and lubricant. Waste treatment was determined from costs for industrial wastewater advertised by the Lake Charles Public Works Department (31). Utility expenditures were determined using common utility costs in the Gulf Coast region. Operating labor was calculated using Equation 36 to determine the number of operators required and an hourly rate of $29.00 per hour for each operator working 2,080 hours per year. For the proposed process, fourteen operators will be hired and each will be paid $60,320 annually.

$$N_{OL}=(6.29+31.7P^2+.23N_{np})^{.5} \quad 36$$

Table 52: Variables for Operating Labor Equation

Symbol	Designation
N_{OL}	**Number of Operators**
N_{np}	**Number of Processes**
P	**People Handling Particulate Solids**

Administration and engineering, maintenance and repairs, operating supplies, laboratory charges, patents and royalties are accounted into the direct costs for the process using proportions provided in Turton. Fixed costs included local taxes, insurance, and plant overhead and were calculated using percentages of fixed capital and cost of manufacturing. General expenses were calculated in a similar manner and included administration costs, distribution and sales, and research and development. Table 52 displays the values for total annual costs, while Figure 12 shows the percentages of each category.

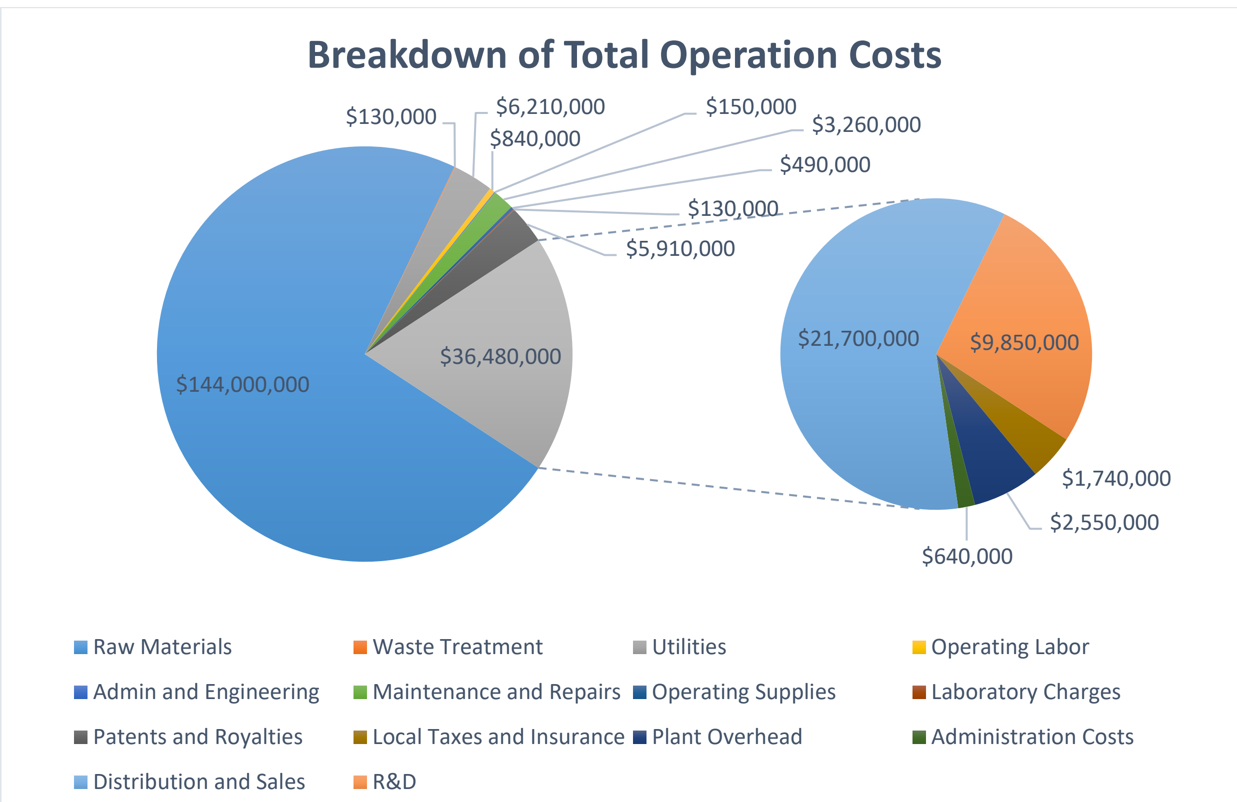


Figure 12: Pie Chart of Annual Costs. The primary circle represents the total operation costs, while the secondary chart displays the breakdown of general expenses and fixed costs.

Table 53: Annual Cost Breakdown and Determination Method

Cost	Value	Method of Determination
Fixed Capital Investment	$ 54,300,000	Total CTM of Project
Cost of Manufacturing	$ 197,000,000	18% of Fixed Capital Income + 2.73*Operating Labor+1.23*(Raw Materials +Utilities + Waste Treatment)
Direct Costs	$ 161,200,000	Raw Materials + Utilities + Waste Treatment +1.33* Operating Labor + 3% of Costs of Manufacturing + 6.9% of Fixed Capital Investment
Raw Materials	$ 144,000,000	Methanol + Lubricant
Waste Treatment	$ 130,000	Calculated from Lake Charles WWTP
Utilities	$ 6,210,000	Steam, Cooling Water, and Electricity
Operating Labor	$ 840,000	14 operators at $29/hr
Admin and Engineering	$ 150,000	18% of Operating Labor
Maintenance and Repairs	$ 3,260,000	6% of Fixed Capital Investment
Operating Supplies	$ 490,000	.9% of Fixed Capital Investment
Laboratory Charges	$ 130,000	15% of Operating Labor
Patents and Royalties	$ 5,910,000	3% of Cost of Manufacturing
Fixed Costs	$ 4,290,000	70.8% of Operating Labor + 6.8% of Fixed Capital Investment
Depreciation	$ -	Covered Separately with MACRS
Local Taxes and Insurance	$ 1,740,000	3.2% of Fixed Capital Investment
Plant Overhead	$ 2,550,000	70.8% of Operating Labor + 3.6% of Fixed Capital Investment
General Expenses	$ 32,160,000	17.7% of Operating Labor + .9% of Fixed Capital Investment + 16% of Cost of Manufacturing
Administration Costs	$ 640,000	17.7% of Operating Labor + .9% of Fixed Capital Investment
Distribution and Sales	$ 21,700,000	11% of Cost of Manufacturing
R&D	$ 9,850,000	5% of Cost of Manufacturing
Total Costs	$ 197,600,000	Raw Materials + Utilities + Waste Treatment +2.215*Operating Labor + 19% of Cost of Manufacture + 14.6% of Fixed Capital Investment

Economic Analysis

The project was analyzed using a hurdle rate of 8% and a 20-year project life. Project expenditures were spread out from 2018 to 2022 using a 5-year disbursement schedule. The expense schedule is estimated below and accounts for preliminary design, detailed engineering, plant construction and permitting, and commissioning. The disbursement schedule is displayed in Table 53.

Table 54: Capital Disbursement Schedule

Year	Percent of Capital Cost Spent
2018	**6%**
2019	**26%**
2020	**20%**
2021	**28%**
2022	**20%**

The process plant qualifies for standard 10-year MACRS depreciation under IRS Publication 946. Total capital expended was depreciated starting in 2023 using the half-year convention. The 20-year plant analysis starts in 2023 and continues until 2042. Working capital was calculated as the cost to fill the methanol storage tanks plus 15% of the fixed capital investment. Working capital was assumed to be spent in 2022 and was recovered at the end of 2042.

NPV and ROR Calculations

The calculated capital and operating costs were incorporated into a spreadsheet to determine the economic attractiveness of the project. For each year that the plant operates, DME revenue was calculated by multiplying the annual production by the sales price per gallon. Direct costs, general expenses, depreciation, and fixed costs were deducted from the DME revenue to find the taxable income. A tax rate of 40% was deducted to produce the net income. Depreciation and other accounting expenses were added back in, and both fixed and working capital were deducted to produce the annual cash flow.

The cash flow was brought back to 2018 using a P/F discount factor for each year, and the net present value (NPV) was determined by summing the discounted cash flows. The discounted cash flow rate of return (DCFROR) was found by applying the Internal Rate of Return (IRR) function in Excel to the cash flow. Working capital was written off at the end of the plant lifespan, but salvage value was assumed to be negligible. This was assumed to be a loss forward situation, with tax benefits deferred to the next year and tax benefits not being realized in the final year of the project.

Washout was assumed throughout the analysis, in which any increases in operating costs are passed directly on to the consumer. This is a reasonable assumption, as the price of methanol follows similar trends to that of diesel, which DME will be competing with in the marketplace. As the price of methanol feed increases, the market price of DME will also increase, allowing profit to remain constant. If the DME price drops, the methanol feed price will also drop.

The project charter stated that the required hurdle rate is 8%. To overcome the discount rate, the sale price of DME per gallon must be \$2.26. At this price, the NPV for the project is \$2.5 million and the DCFROR is 8.52%. As the DCFROR is greater than 8% and the NPV is greater than zero, the project is economically attractive at these conditions. An abbreviated cash flow table is shown below, while the complete cash flow table can be found in Appendix C: Cash Flow Table.

Project Title:	*Propylene Splitter*		Fixed Cap	$ 54,313,228	
Corporate financial situation	*Expense*		Cost of MeOH/gal	$2.26	
Corporate Rate of Return i*	8%		Operating Cost	$ 6,232,445	
Other Relevant Project Info	*10 Year MACRS Depreciation*		Working Capital	$ 15,646,984	
End of Year	2018	2022	2023	2024	2042
Production DME (U.S Gallons/year)			91,250,000	91,250,000	91,250,000
x Sales Price ($/U.S Gallon)			$ 2.26	$ 2.26	$ 2.26
+Net Revenue			$ 206,225,000	$ 206,225,000	$ 206,225,000
- Raw Material Costs (Lubricant+ Methanol)			$ (141,600,519)	$ (141,600,519)	$ (141,600,519)
Methanol (U.S Gallons/year)			94,224,278	94,224,278	94,224,278
x cost Price ($/U.S Gallons)			1.49	1.49	1.49
x cost Shipment cost ($/U.S Gallon)			0.02	0.02	0.02
Lubrication (lb/year)			410,143.20	410,143.20	410,143.20
Lubrication Cost ($/lb)			1.65	1.65	1.65
Lubrication (U.S Gallons/year)			70208	70208	70208
Lubrication Shipping Cost ($/U.S Gallon)			0.02	0.02	0.02
Total Lubrication Cost $/year			678,140.44	678,140.44	678,140.44
Waste Water Treament (U.S Gallons/year)			29,807,652	29,807,652	29,807,652
- Waste Water Treament ($/year)			$ (131,750)	$ (131,750)	$ (131,750)
- Other Direct Costs			$ (9,937,528)	$ (9,937,528)	$ (9,937,528)
- Fixed Costs			$ (4,291,191)	$ (4,291,191)	$ (4,291,191)
- General Expenses			$ (32,164,889)	$ (32,164,889)	$ (32,164,889)
- Utilities			$ (6,232,445)	$ (6,232,445)	$ (6,232,445)
- Manual Labor Costs			$ (844,480)	$ (844,480)	$ (844,480)
- Depreciation			$ (5,431,323)	$ (9,776,381)	$ -
Depreciation Ratio, 10-year MACRS			0.1000	0.1800	
- Loss Forward			$ -	$ -	$ -
- Write Off					$ (15,646,984)
Taxable Income			$ 5,590,874	$ 1,245,816	$ (4,624,787)
- Tax @ 40%			$ (2,236,350)	$ (498,326)	$ -
Net Income			$ 3,354,525	$ 747,490	$ (4,624,787)
+ Depreciation			$ 5,431,323	$ 9,776,381	$ -
+ Loss Forward			$ -	$ -	$ -
+ Write Off			$ -	$ -	$ 15,646,984
- Working Capital		$ (15,646,984)	$ -	$ -	$ -
- Fixed Capital equipment cost	$ (3,258,794)	$ (10,862,646)	$ -	$ -	$ -
Fixed Capital expense schedule	0.06	0.20	0	0	0
Cash Flow	$ (3,258,794)	$ (26,509,629)	$ 8,785,847	$ 10,523,871	$ 11,022,197
Discount Factor(P/F_i',n)	1.0000	0.7350	0.6806	0.6302	0.1577
Discounted Cash Flow	$ (3,258,794)	$ (19,485,369)	$ 5,979,500	$ 6,631,824	$ 1,738,193
NPV @i' =	$ 2,488,958	$ 2,500,000			
DCFROR =	8.52%	8.5%			

Figure 13: Abbreviated Cash Flow Table.

Though these conditions suggest that the project is economically attractive, it is based strongly on the sale price of DME. To be competitive in the market, DME should be priced at around half the cost of diesel fuel, which at this time would be approximately $1.50. As this price is even less than the cost of equivalent feedstock methanol, it is not possible to market DME at that price. Instead, this analysis shows that DME must be sold for 51% more than an energy equivalent of diesel. Incentives to motivate adoption are discussed later in the economic analysis section.

Net cash flow is displayed in Figure 14, while net project value is displayed in Figure 15. When pure cash flow is considered, the initial capital investment is recovered shortly through the year 2020, whereas the project only turns a profit by 2041 when the time value of money is accounted for. The first five years of the project show a steep decrease in project value as the process is constructed and no revenue is generated. After 2022, the project starts to recover value as profit is generated and depreciation starts to take affect. After the ten years of depreciation run out, the project recovers value at a constant rate until writeoff of working capital produces a spike in the final year.

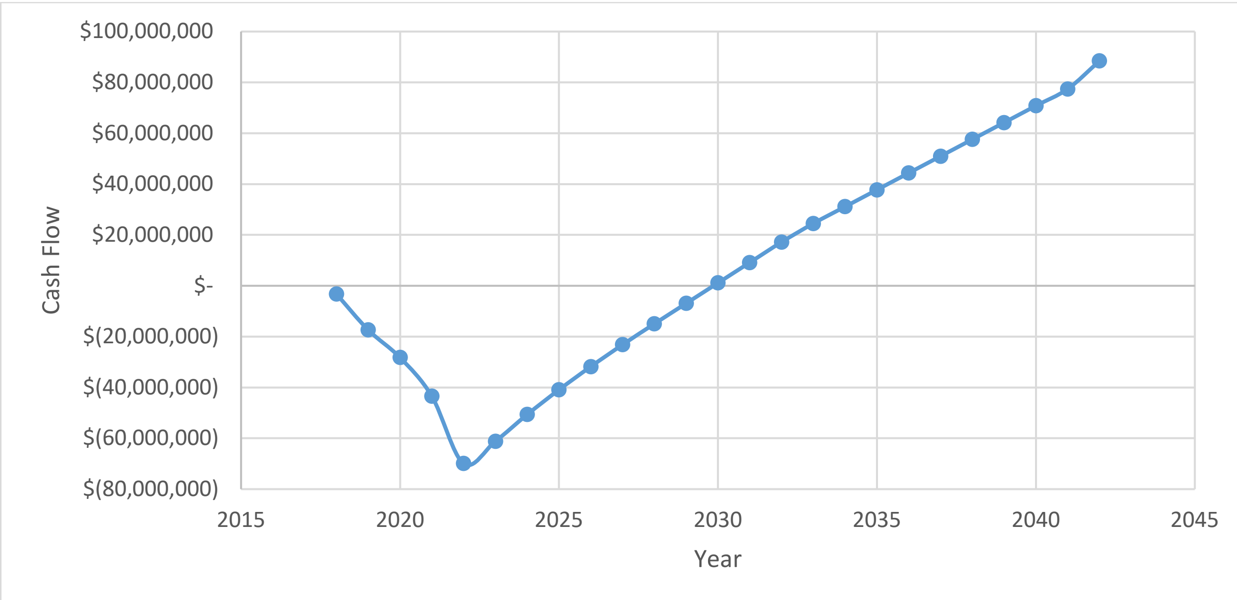


Figure 14: Non-discounted Cash Flow Diagram.

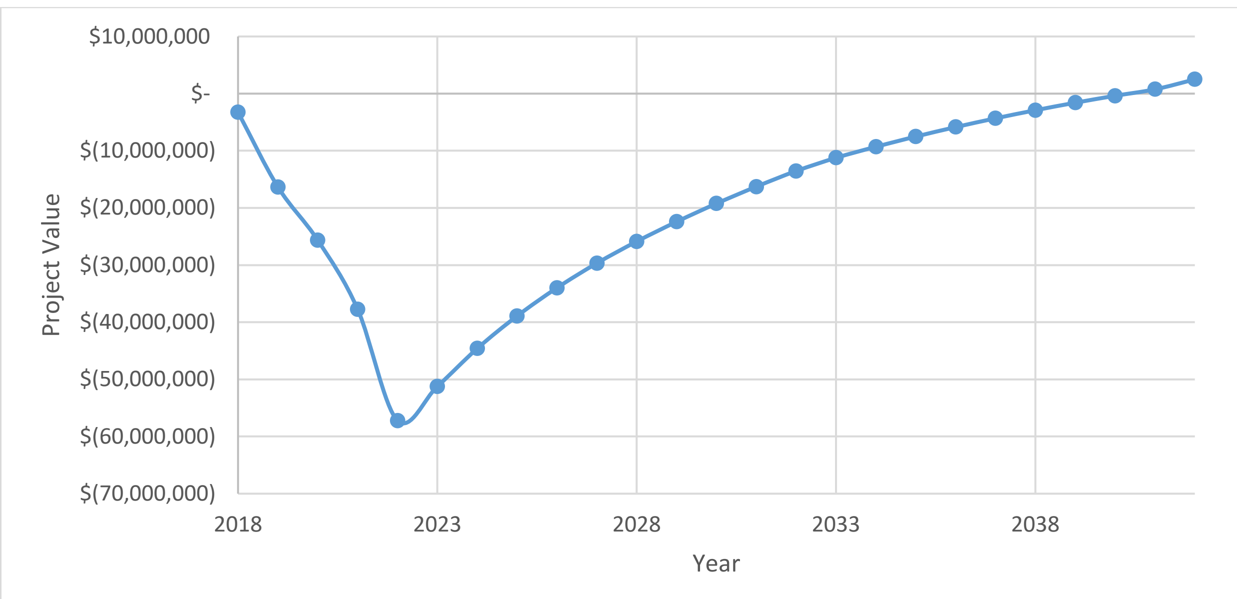


Figure 15: Project value vs time with a sales price of $2.26/gal. Accounts for the time value of money.

Sensitivity Analysis

The tornado charts in Figure 16 and Figure 17 describe the variability in the economic attraction of the project. DME sales price was by far the most sensitive value, as a 5% drop in the sales price of DME led to a DCFROR of -7.42% and an NPV of -$41.6 million. A 5% increase in the DME sales price did increase the profitability of the project, but not as significantly as the negative impact the 5% decrease had on the metrics. After DME sales price, a raw material cost variation of 5% had the next greatest impact on the ROR and NPV, followed subsequently by the fixed capital investment and the transportation cost.

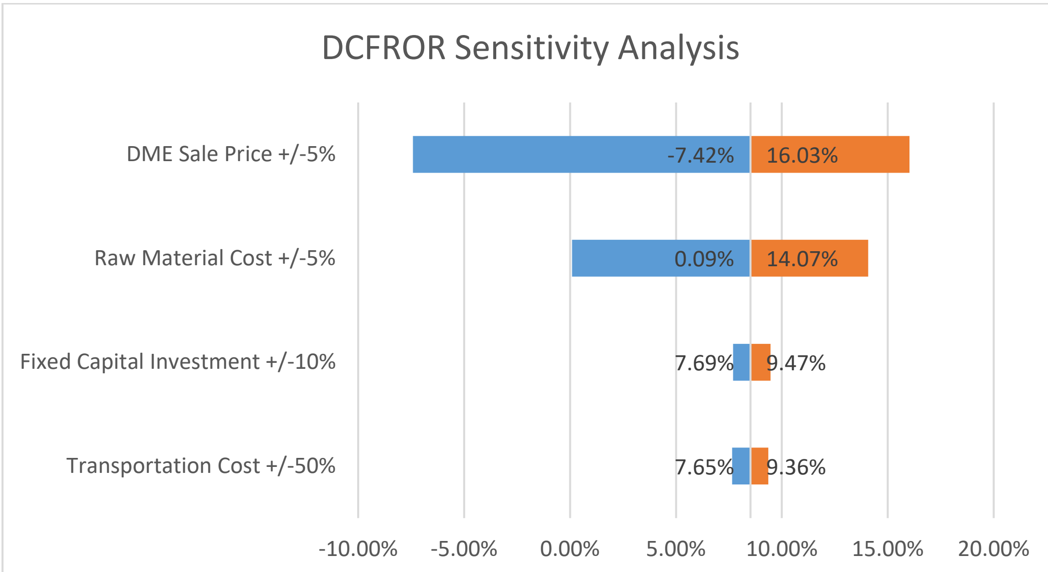


Figure 16: Tornado Chart for DCFROR

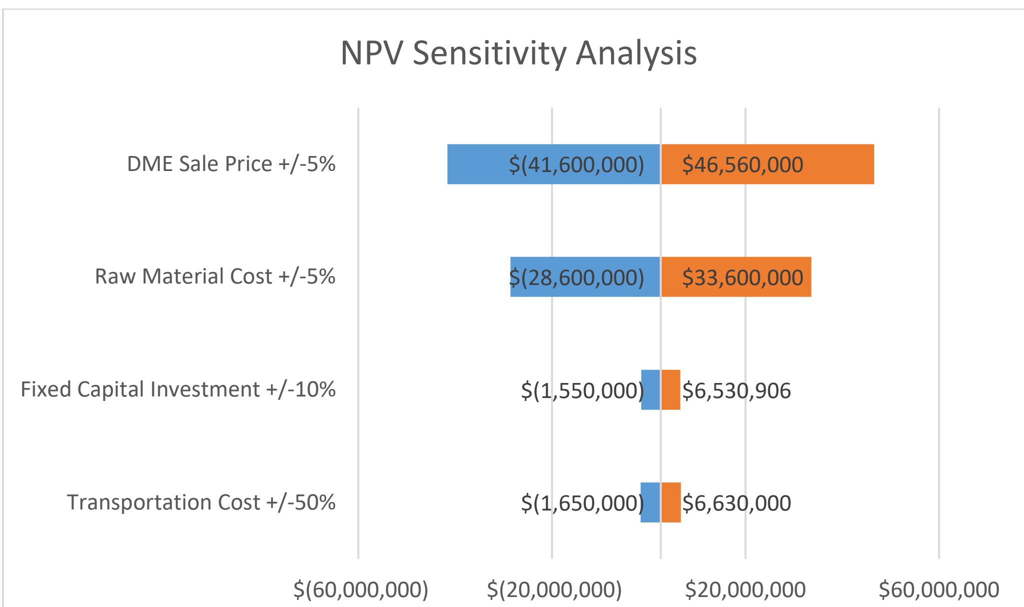


Figure 17: Tornado Chart for NPV

Table 55: Best and Worst Case Scenario

	Best Case Scenario	Worst Case Scenario
ROR	21.45%	N/A
NPV	$ 83,600,000	$ (81,100,000)

The best and worst case scenarios were determined through combining the benefits and drawbacks of the previously considered variations. In the best case, the DCFROR becomes greater than 20% and the NPV exceeds $80 million. In the worse-case, however, the NPV is below negative $80 million and the ROR was so low that the embedded Excel function IRR was unable to solve. The project is highly volatile and even in the best reasonable scenario the project nets less than $100 million over 20-years, whereas the worst-case scenario results in significant economic losses.

From an economic perspective, the project is a risky investment. To reach the point where the project is economically attractive, the DME must be sold at a point that is 51% more than an energy equivalent of gasoline. The risk analysis assumes that this price can be achieved and the variation of the economics was based around this value calculated to overcome the discount rate. If this value is missed by even a penny per gallon, the project will fail to meet the required rate of return. As such, much caution is required when considering this project, as many things must go right for the project to turn a profit, while only a few things need to change to ruin its economic viability.

Economic Incentives

The risk on this project can be reduced by taking advantage of the available government subsidies designed to promote the use of clean, renewable fuels. Many of the economic incentives provided by various government agencies are focused on the development of biodiesel, ethanol, and liquefied petroleum gas, with few towards DME promotion. There are some economic incentives directed toward DME transportation, including Louisiana's 30% tax credit toward converting diesel trucks to DME, but they are ultimately insufficient to encourage mass adoption of DME. One program that has the potential to increase the profitability of DME fuel synthesis is the EPA Renewable Fuel Standard Program. Within this program, companies that generate fuel from renewable sources qualify for RINs, which can then be sold to companies that accrue deficits from producing non-renewable fuels.

As this is a lucrative means of income that could highly affect the economic viability of the process, it is essential that the proposed DME plant qualify for the program. The methanol that is used for feedstock will be received from the new Lake Charles methanol plant, which is designed to refine waste petroleum coke into methanol. Though this feed comes from a non-renewable resource, it does take a waste stream and converts it into a clean feedstock for a number of processes. If this source qualifies for the RIN program, the proposed DME process would be able to sell the product at lower than the currently required price of $2.26/gal. If the current source of methanol does not qualify for the program, it is suggested that the plant be relocated to a location where a renewable source of methanol is more readily available.

As legislation and regulation advances in the upcoming decades, the incentives to adopt DME as a fuel alternative will likely increase. In addition, the incentives currently put towards biodiesel and ethanol may be expanded to include DME. Despite these possibilities, neither future government subsidies nor current RIN sales were included in the economic analysis to ensure a measure of conservatism in the overall analysis.

Conclusions and Recommendations

The proposed design utilizes a high-temperature vapor phase reaction over a gamma-alumina catalyst to produce fuel-grade DME and water from methanol. This process was selected because of its proven industrial track record, low fixed capital costs, and ease of operation and control. The process was split in half to allow for efficient 50% turndown while operating at design operating conditions, to lower the fixed cost as compared to single unit operations, and to maximize plant flexibility for maintenance and operability.

The process utilizes high pressure pumps to send liquid methanol through a series of heat exchangers and a furnace for preheating. The preheated stream flows through the reactors and exits at 70% conversion. The heat from this stream is integrated into the reactant stream and the outlet mixture is then fed to a distillation unit. The first distillation column removes high-purity DME off the top, from where it is mixed with lubricant and sent to the truck filling station. The bottom stream of water and methanol is fed to a second distillation unit, where the methanol is recovered and sent to recycle, while the wastewater is purified below the lower flammability limit of methanol, cooled, and discharged to the sanitary sewer for treatment.

Based on a sale price of $2.26/gallon of DME and a washout assumption, the NPV for the project is $2.5 million and the DCFROR is 8.52%, which overcomes the discount rate of 8%. At these conditions, the initial investment of $70 million is recovered by 2030 when considering non-discounted cash flow, but the project only gains net value by the 2041 when accounting for the time value of money.

Sensitivity analysis shows that the economic parameters are most highly affected by the sale price of DME followed by the purchase price of reactants and then the fixed capital investment. To be economically viable, the DME must be sold for $2.26/gallon, whereas if it drops by even 5% the project ROR falls below -7%. In the unregulated market, DME must be sold at a competitive energy equivalent price to that of diesel, but for this project to be economically attractive the sale price of DME must be at least 51% higher than its diesel energy equivalent.

The risk and market analyses suggest that the likelihood of the project returning value on its own are low. There are a few government programs that are designed to incentivize the adoption and production of renewable and clean energy sources, including the EPA Renewable Fuel Standard and the Louisiana Fueling Infrastructure Tax Credit, but their application and overall effect were not accounted for in the economic analysis.

This process has the overall benefit of reducing the carbon footprint of trucks operating in the vicinity of the plant in comparison to diesel. Due to this, it does have an overall reduced carbon footprint in comparison to the alternatives. To further reduce the footprint, an alternative fuel could be burned in the furnace and all equipment in the plant could be converted to DME fuel. In addition, all process utilities should be designed to utilize clean energy sources to ensure the smallest carbon footprint possible.

The wastewater stream was utilized for monetary compensation. Initial designs had higher concentrations of methanol going to the wastewater plant, but the feedstock lost was a significant expense for the process. The process was redesigned with a higher purity of water specified. As this wastewater contains very low quantities of methanol, the amount of wasted methanol is minimized. This required a larger separation unit, but it ultimately resulted in net profits as the vast majority of methanol is recycled to the head of the process.

Without the assistance of government incentives, it is not recommended to pursue this project to the detailed design phase. The price point for DME must be set higher than is reasonably competitive with diesel, and the overall economic attractiveness is highly dependent on this parameter. As such, economic success is dependent on qualification for and utilization of all possible means of secondary income, including transportation programs, environmental initiatives, and clean energy incentives. It is recommended that further research into these programs be the primary concern if the project is moved into detailed design.

Acknowledgements

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Appendix

Appendix A: Additional Reactor Calculations

POLYMATH Report No Title
Ordinary Differential Equations 05-Mar-2018

Calculated values of DEQ variables

	Variable	Initial value	Minimal value	Maximal value	Final value
1	A1	0.5366	0.5366	0.5366	0.5366
2	A2	0.045	0.045	0.045	0.045
3	Afor	1.063E+06	1.063E+06	1.063E+06	1.063E+06
4	alpha	0.069198	0.069198	0.069198	0.069198
5	Arev	1.468E+07	1.468E+07	1.468E+07	1.468E+07
6	Ca	0.73471	0.167614	0.73471	0.167614
7	Cao	0.73471	0.73471	0.73471	0.73471
8	Cb	0	0	0.2040769	0.2040769
9	Cbo	0.000222	0.000222	0.000222	0.000222
10	Cc	0	0	0.2040769	0.2040769
11	Cco	0.01469	0.01469	0.01469	0.01469
12	Cpa	70.	70.	70.	70.
13	Cpb	110.	110.	110.	110.
14	Cpc	36.85	36.85	36.85	36.85
15	DeltaCp	3.425	3.425	3.425	3.425
16	E1	-3450.	-3450.	-3450.	-3450.
17	E2	-9395.	-9395.	-9395.	-9395.
18	Efor	6.563E+04	6.563E+04	6.563E+04	6.563E+04
19	Erev	8.899E+04	8.899E+04	8.899E+04	8.899E+04
20	f1c	0	0	0	0
21	f2c	155.	155.	155.	155.
22	fa	0.5	0.5	0.5	0.5
23	Fao	0.103	0.103	0.103	0.103
24	Hrxn	-1.171E+04	-1.171E+04	-1.171E+04	-1.171E+04
25	K1	1.136419	1.010156	1.136419	1.010156
26	K2	0.3472695	0.2519856	0.3472695	0.2519856

27	kfor	0.6709022	0.6709022	6.305786	6.305786
28	krev	0.0575759	0.0575759	1.201358	1.201358
29	m	2.	2.	2.	2.
30	n	4.	4.	4.	4.
31	R	8.314	8.314	8.314	8.314
32	ra	-0.0238467	-0.0526328	-0.0238467	-0.0318394
33	SumThetaCp	72.19955	72.19955	72.19955	72.19955
34	T	553.	553.	655.9561	655.9561
35	To	553.	553.	553.	553.
36	W	0	0	1.81	1.81
37	X	0	0	0.7088862	0.7088862
38	y	1.	0.9295678	1.	0.9295678

Differential equations

1 d(X)/d(W) = -ra/Fao

2 d(y)/d(W) = -alpha/(2*y)*(T/To)

Explicit equations

1 alpha = .069198

2 To = 553

Kelvin

3 Hrxn = -11712

kJ/kgmole, assume at 298K

4 Afor = 1.0626*(10^6)

kgmole/m^3*s

5 Efor = 65633.0

kJ/kgmole

6 Cpc = 36.85

kJ/kmol K

7 Arev = 1.4677*(10^7)

kgmole/m^3*s

8 Erev = 88994.0

KJ/kgmole

9 Cpb = 110

```
kJ/kmol K
Cpa = 70
kJ/Kgmol*K
Cao = .73471
kgmole/m^3
m = 2
n = 4
DeltaCp = .5*(Cpb + Cpc) - Cpa
A1 = 0.5366
E1 = -3450.0
kJ/kgmole
SumThetaCp = Cpa + .0196/.9802* Cpb
A2 = 0.045
E2 = -9395.0
kJ/kgmole
fa = 0.5
f1c = 0
f2c = 155
R = 8.314
kJ/(kgmole*K)
T = (X*(-Hrxn)+SumThetaCp*To + X*DeltaCp*298)/(SumThetaCp + X*DeltaCp)
K2 = A2*exp(-E2/(R*T))
*((1/T2) - (1/T)))
kfor = Afor*exp(-Efor/(R*T))
*((1/T1)-(1/T)))
krev = Arev*exp(-Erev/(R*T))
*((1/T1)-(1/T)))
Ca = Cao*(1-X)*(To/T)*y
Cb = Cao*(.5)*X*(To/T)*y
K1 = A1*exp(-E1/(R*T))
*((1/T2)-(1/T))
Fao = .103
kgmole/s
Cc = Cao*(.5)*X*(To/T)*y
```

33 Cbo = .000222

kgmole/m^3

34 Cco = 0.01469

kgmole/m^3

35 ra = -((kfor*(Ca^m)) - (krev*Cb*Cc))/((1+(K1*(Ca^fa)*(Cc^f1c))+(K2*(Cc^f2c)))^n)

General

Total number of equations	37
Number of differential equations	2
Number of explicit equations	35
Elapsed time	1.157 sec
Solution method	RKF_45
Step size guess. h	0.000001
Truncation error tolerance. eps	0.000001

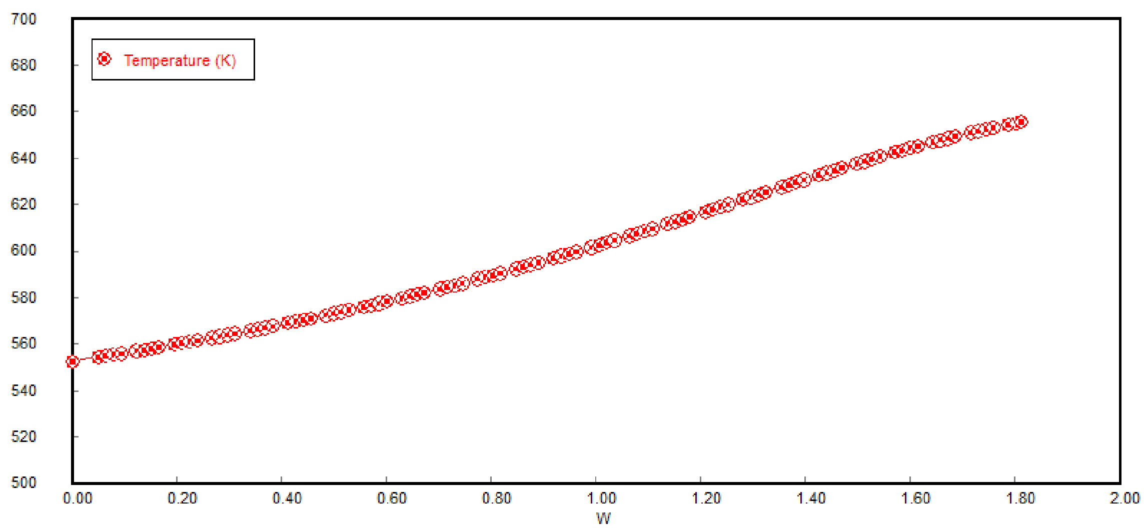


Figure 18: Temperature (K) vs. Catalyst Volume (m^3)

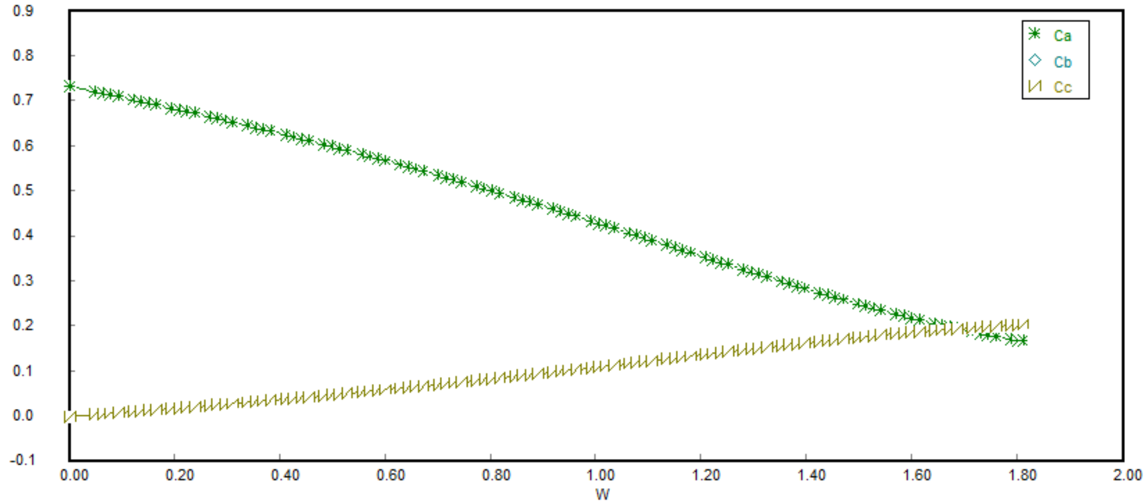


Figure 19: Reactant and Product Concentrations (kmol/m^3) vs. Catalyst Volume (m^3)

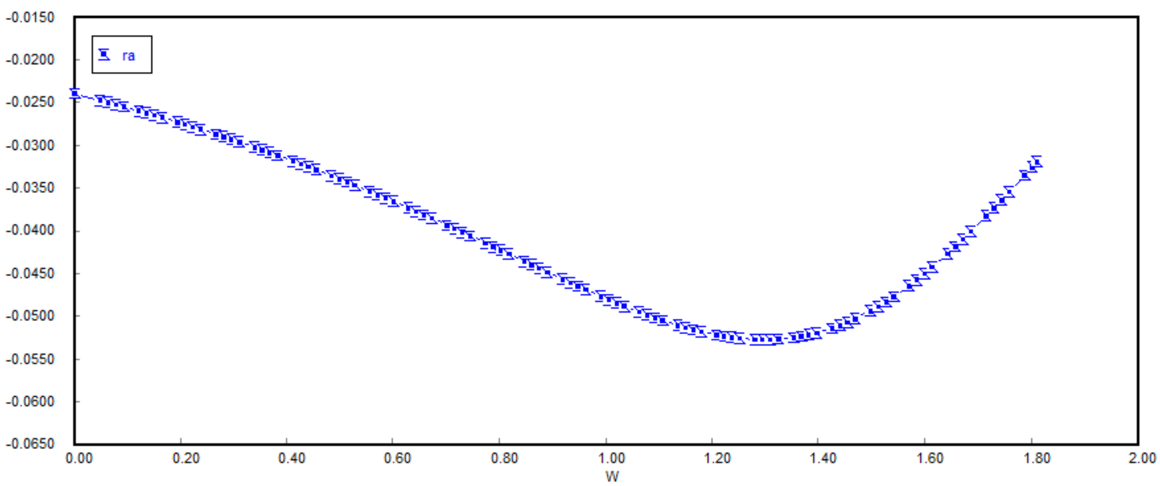


Figure 20: Methanol reaction rate (kmol/m^3s) vs. Catalyst Volume (m^3)

Appendix B: HEX Calculations

E-101

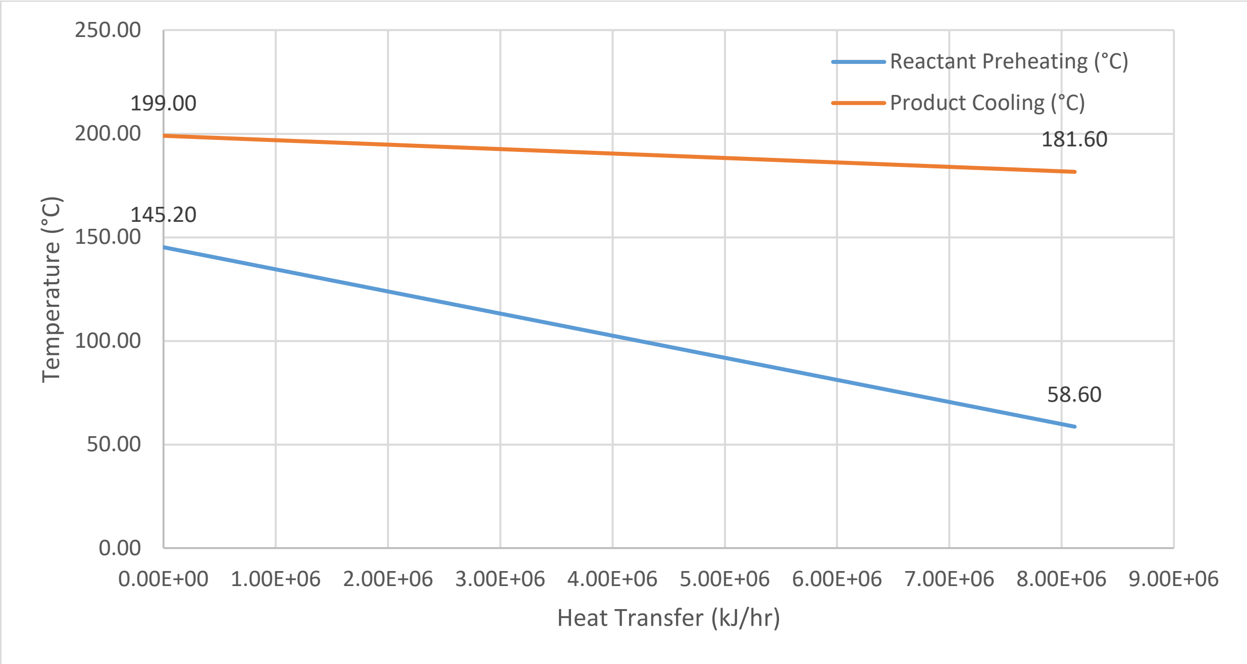


Figure 21: T-Q Diagram for E-101 A/B

Table 56: HEX Calculation Table for E-101

Q (kJ/h)	Reactant Preheating (°C)	Product Cooling (°C)	Delta T
9.34E+06	145.2	199	53.8
1.75E+07	58.6	181.6	123
8.12E+06		LMTD	83.7
Function	Heating liquid	Condensing vapor	
HEX Side	Tube Side	Shell Side	

Variable	Value	Units
Q	2254.5	kW
Uo=	652.17	W/(m^2)*K
F	0.9751	2 shell pass, 4 tube pass
LMTD	83.7	°C

Variable	Value	Units
hi	1500	W/(m^2)*K
Rfi	0.0001	((m^2)*K)/W
ho	1500	W/(m^2)*K
Rfo	0.0001	((m^2)*K)/W

A=	42.3	m^2
A=	456	ft^2
A*1.1=	46.6	m^2
A*1.1=	502	ft^2

Table 57: Cost Calculations for E-101 A/B

Heat Exchanger Costing Calculator						
Variable	Equation	U-Tube	Floating Head	Fixed Tube	Bayonet	Units
C_{po} Pg. 910	$\log_{10}(C_P^0)=K_1+K_2\log_{10}(A)+K3\log_{10}(A)]^2$	$ 20,905.81	$ 19,864.38	$ 18,793.35	$ 39,132.19	$
K_1	Table A.1 in Turton	4.1884	4.8306	4.3247	4.2768	N/A
K_2	Table A.1 in Turton	-0.2503	-0.8509	-0.303	-0.0495	N/A
K_3	Table A.1 in Turton	0.1974	0.3187	0.1634	0.1431	N/A
Finding F_{BM}	FBM=B1+B2*FM*FP	3.57	3.57	3.57	3.57	N/A
Table A.4 pg. 934: B1	B1	1.63	1.63	1.63	1.63	N/A
Table A.4 pg. 935: B2	B2	1.66	1.66	1.66	1.66	N/A
Pg. 931 F_M	CS Shell and Tube	1	1	1	1	N/A
Fp (A.3)	og10(FP)=C1+C2log10(P)+(C3((log10(P))^2	1.17	1.17	1.17	1.17	N/A
Pressure	Converting psia to barg	40.00	40.00	40.00	40.00	barg
C1		0.03881	0.03881	0.03881	0.03881	$
C2		-0.11272	-0.11272	-0.11272	-0.11272	N/A
C3		0.08183	0.08183	0.08183	0.08183	N/A
$C_{BM\ 2001}$	$C_{BM}=Cp^o*F_{BM}$	$ 74,685.77	$ 70,965.26	$ 67,139.02	$ 139,799.31	$
CAPCOST	CEPCI 2001: 397					N/A
	CEPCI 2017: 566.6					N/A
$C_{BM\ 2017}$	CBM2017=CBM2001*(566.6/397)	$ 106,591.83	$ 101,281.90	$ 95,821.08	$ 199,522.14	$
Contingency	CBM2017*0.15	$ 15,988.77	$ 15,192.29	$ 14,373.16	$ 29,928.32	$
Fees	CBM2017*0.03	$ 3,197.75	$ 3,038.46	$ 2,874.63	$ 5,985.66	$
C_{TM}		$ 125,778.36	$ 119,512.65	$ 113,068.88	$ 235,436.12	$
Rounded		$ 126,000.00	$ 120,000.00	$ 113,000.00	$ 235,000.00	
For both HEX		$ 252,000.00	$ 240,000.00	$ 226,000.00	$ 470,000.00	

E-102

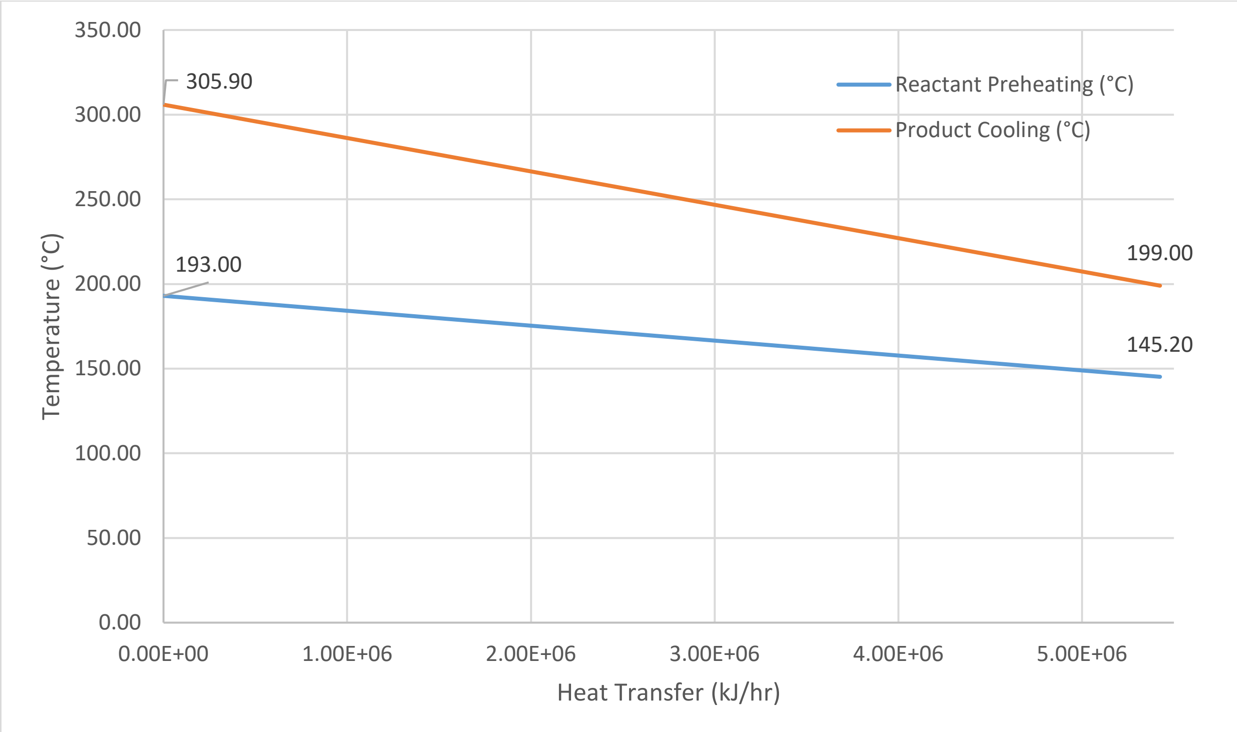


Figure 22: T-Q Diagram for E-102 A/B

Table 58: HEX Calculation Table for E-102

Q (kJ/h)	Reactant Preheating (°C)	Product Cooling (°C)	Delta T
3.92E+06	**193**	**305.9**	**112.9**
9.34E+06	**145.2**	**199**	**53.8**
5.42E+06		**LMTD**	**79.7**
Function	**Heating liquid**	**Cooling gas**	
HEX Side	**Tube Side**	**Shell Side**	

Variable	Value	Units
Q	**1506.75**	**kW**
Uo=	**415.78**	**W/(m^2)*K**
F	**0.9751**	**2 shell pass, 4 tube pass**
LMTD	**79.7**	**°C**

Variable	Value	Units
hi	**1500**	**W/(m^2)*K**
Rfi	**0.0001**	**((m^2)*K)/W**
ho	**650**	**W/(m^2)*K**
Rfo	**0.0001**	**((m^2)*K)/W**

A=	46.6	m^2
A=	502	ft^2
A*1.1=	51.3	m^2
A*1.1=	552	ft^2

Table 59: Cost Calculations for E-102 A/B

Heat Exchanger Costing Calculator						
Variable	Equation	U-Tube	Floating Head	Fixed Tube	Bayonet	Units
C_{po} Pg. 910	$\log_{10}(C_P^0)=K_1+K_2\log_{10}(A)+K3\log_{10}(A)]^2$	$ 21,754.90	$ 20,297.62	$ 19,245.95	$ 40,789.16	$
K_1	Table A.1 in Turton	4.1884	4.8306	4.3247	4.2768	N/A
K_2	Table A.1 in Turton	-0.2503	-0.8509	-0.303	-0.0495	N/A
K_3	Table A.1 in Turton	0.1974	0.3187	0.1634	0.1431	N/A
Finding F_{BM}	FBM=B1+B2*FM*FP	3.57	3.57	3.57	3.57	N/A
Table A.4 pg. 934: B1	B1	1.63	1.63	1.63	1.63	N/A
Table A.4 pg. 935: B2	B2	1.66	1.66	1.66	1.66	N/A
Pg. 931 F_M	CS Shell and Tube	1	1	1	1	N/A
Fp (A.3)	log10(FP)=C1+C2log10(P)+(C3((log10(P))^2)	1.17	1.17	1.17	1.17	N/A
Pressure	Converting psia to barg	40.00	40.00	40.00	40.00	barg
C1		0.03881	0.03881	0.03881	0.03881	$
C2		-0.11272	-0.11272	-0.11272	-0.11272	N/A
C3		0.08183	0.08183	0.08183	0.08183	N/A
$C_{BM\ 2001}$	$C_{BM}=Cp^o*F_{BM}$	$ 77,719.12	$ 72,513.00	$ 68,755.95	$145,718.80	$
CAPCOST	CEPCI 2001: 397					N/A
	CEPCI 2017: 566.6					N/A
$C_{BM\ 2017}$	CBM2017=CBM2001*(566.6/397)	$110,921.04	$ 103,490.85	$ 98,128.77	$207,970.45	$
Contingency	CBM2017*0.15	$ 16,638.16	$ 15,523.63	$ 14,719.32	$ 31,195.57	$
Fees	CBM2017*0.03	$ 3,327.63	$ 3,104.73	$ 2,943.86	$ 6,239.11	$
C_{TM}		$130,886.83	$ 122,119.21	$115,791.95	$245,405.14	$
Rounded		$131,000.00	$ 122,000.00	$116,000.00	$245,000.00	
For both HEX		$262,000.00	$ 244,000.00	$232,000.00	$490,000.00	

E-104

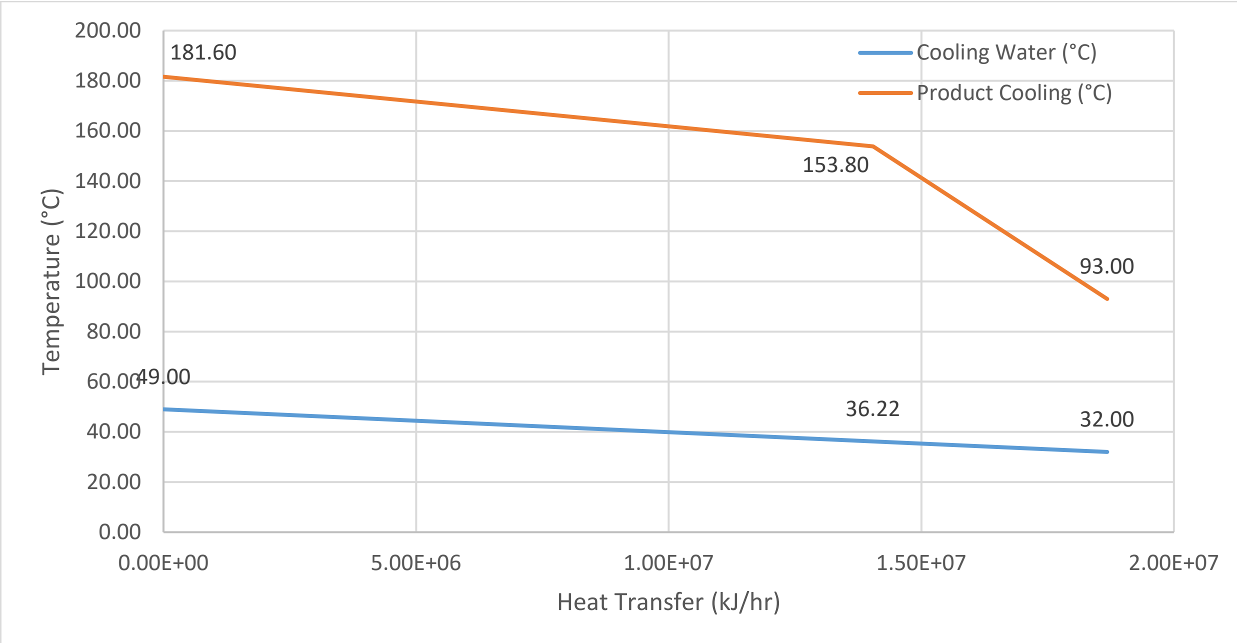


Figure 23: T-Q Diagram for E-104 A/B

Table 60: HEX Calculation Table for E-104

	Tube Side	Shell Side	
Q (kJ/h)	Cooling Water (°C)	Product Cooling (°C)	Delta T
1.75E+07	**49.0**	**181.6**	**132.6**
3.61E+07	**32.0**	**93.0**	**61**
1.87E+07		**LMTD**	**92.2**

Variable	Value	Units
Q	**5189**	**kW**
Uo=	**857**	**W/(m^2)*K**
F	**0.9751**	**2 shell pass, 4 tube pass**
LMTD (K)	**92.2**	**°C**

Variable	Value	Units
hi	**5000**	**W/(m^2)*K**
Rfi	**0.0002**	**((m^2)*K)/W**
ho	**1500**	**W/(m^2)*K**
Rfo	**0.0001**	**((m^2)*K)/W**

A=	67.3	m^2
A=	724.7	ft^2
A*1.1=	74.1	m^2
A*1.1=	797.2	ft^2

Cp Water	75.4	kJ/kmol*K
CW Flow Rate	2688	kmol/hr
CW Flow Rate	48,380	kg/hr

CW Properties		
Cp Water	75.4	kJ/kmol*K
Temp In	32	°C
Temp Out	49	°C
Service Factor	0.95	
Cooling Water Density	993	kg/m^3
Cooling Water Cost	0.05	$/cubic meter
Operating Seconds per year	3.00E+07	seconds/yr

Table 61: Cost Calculations for E-104

Heat Exchanger Costing Calculator						
Variable	Equation	U-Tube	Floating Head	Fixed Tube	Bayonet	Units
C_{po} Pg. 910	$\log_{10}(C_P^0)=K_1+K_2\log_{10}(A)+K3\log_{10}(A)]^2$	$ 25,728.91	$ 22,580.59	$ 21,347.82	$ 48,354.48	$
K_1	Table A.1 in Turton	4.1884	4.8306	4.3247	4.2768	N/A
K_2	Table A.1 in Turton	-0.2503	-0.8509	-0.303	-0.0495	N/A
K_3	Table A.1 in Turton	0.1974	0.3187	0.1634	0.1431	N/A
Finding F_{BM}	FBM=B1+B2*FM*FP	3.57	3.57	3.57	3.57	N/A
Table A.4 pg. 934: B1	B1	1.63	1.63	1.63	1.63	N/A
Table A.4 pg. 935: B2	B2	1.66	1.66	1.66	1.66	N/A
Pg. 931 F_M	Carbon Steel shell and tube	1	1	1	1	N/A
Fp (A.3)	$\log_{10}$(FP)=C1+C2log10(P)+(C3((log10(P))^2)	1.17	1.17	1.17	1.17	N/A
Pressure	Converting psia to barg	40.00	40.00	40.00	40.00	barg
C1		0.03881	0.03881	0.03881	0.03881	$
C2		-0.11272	-0.11272	-0.11272	-0.11272	N/A
C3		0.08183	0.08183	0.08183	0.08183	N/A
$C_{BM\ 2001}$	$C_{BM}=Cp^o*F_{BM}$	$ 91,916.24	$ 80,668.91	$ 76,264.86	$ 172,745.82	$
CAPCOST	CEPCI 2001: 397					N/A
	CEPCI 2017: 566.6					N/A
$C_{BM\ 2017}$	CBM2017=CBM2001*(566.6/397)	$ 131,183.23	$ 115,130.99	$ 108,845.51	$ 246,543.53	$
Contingency	CBM2017*0.15	$ 19,677.48	$ 17,269.65	$ 16,326.83	$ 36,981.53	$
Fees	CBM2017*0.03	$ 3,935.50	$ 3,453.93	$ 3,265.37	$ 7,396.31	$
C_{TM}		$ 154,796.21	$ 135,854.57	$ 128,437.70	$ 290,921.36	$
Rounded		$ 155,000.00	$ 136,000.00	$ 128,000.00	$ 291,000.00	
For both HEX		$ 310,000.00	$ 272,000.00	$ 256,000.00	$ 582,000.00	
Operating Cost:						
Heat Transferred		5188.96	kW			
CW energy required		1.55E+11	kJ/yr			
Water flow rate		4.049793661	kmol/s			
Volumetric flow rate		0.07	cubic meters/s			
Water Required per year		2,198,866.70	cubic meters/yr	For both HEX	Rounded	
Annual Cost of CW		$ 109,943.34	$/yr	$ 219,886.67	$ 220,000.00	
PWC of CW @ 8% ROR and 20 yr lifetime		$ 1,079,423.66		$ 2,158,847.33	$ 2,160,000.00	
Total PWC of HEX		$ 1,207,861.37		$ 2,415,722.73	$ 2,420,000.00	

H-101

Table 62: Furnace Calculation Table for H-101

Parameter	Value	Units
Heat duty	3.48E+07	BTU/hr
LHV of Natural Gas (kJ/cubic foot)	896.1	kJ/cubic foot
Service Factor	0.95	
Operating Seconds per Year	3.00E+07	
Cost of Natural Gas ($/cubic foot)	3.5	$/1000 cubic feet

Table 63 Cost Calculations for H-101

A2+A3				
Variable	Equation		Nonreactive Fired Heater	Units
Heat	Q = UAF*LMTD		10161.64	kW
C_{po} Pg. 910	$\log_{10}(C_P^0)=K_1+K_2\log_{10}(A)+K3\log_{10}(A)]^2$	$	851,890.66	$
K_1	Table A.1 in Turton		7.3488	N/A
K_2	Table A.1 in Turton		-1.1666	N/A
K_3	Table A.1 in Turton		0.2028	N/A
Pg. 931 Fb_M	FBM		2.1	N/A
Fp (A.3)	log10(FP)=C1+C2log10(P)+(C3((log10(P))^2)		1.10	N/A
Pressure	Converting psia to barg		40.00	barg
C1			0.1405	$
C2			-0.2698	N/A
C3			0.1293	N/A
$C_{BM\ 2001}$	$C_{BM}=Cp^o*F_{BM}*Fp$	$	1,962,132.15	$
CAPCOST	CEPCI 2001: 397			N/A
	CEPCI 2017: 566.6			N/A
$C_{BM\ 2017}$	CBM2017=CBM2001*(566.6/397)	$	2,800,362.92	$
Contingency	CBM2017*0.15	$	420,054.44	$
Fees	CBM2017*0.03	$	84,010.89	$
C_{TM}		$	3,304,428.24	$
Rounded		$	3,300,000.00	$
Annual kJ Requirement			3.04E+11	kJ/yr
Natural Gas Burned Per Year			3.40E+08	cubic feet/yr
Annual Expense of Fuel		$	1,189,064.48	Annual Cost
Annual Cost, Rounded		$	1,190,000.00	
Present Worth Cost for 20 yrs @ 8% ROR		$	11,674,235.03	PWC
Total Capital and Utility PWC		$	14,978,663.28	
Total Capital and Utility PWC, Rounded		$	15,000,000.00	

E-203

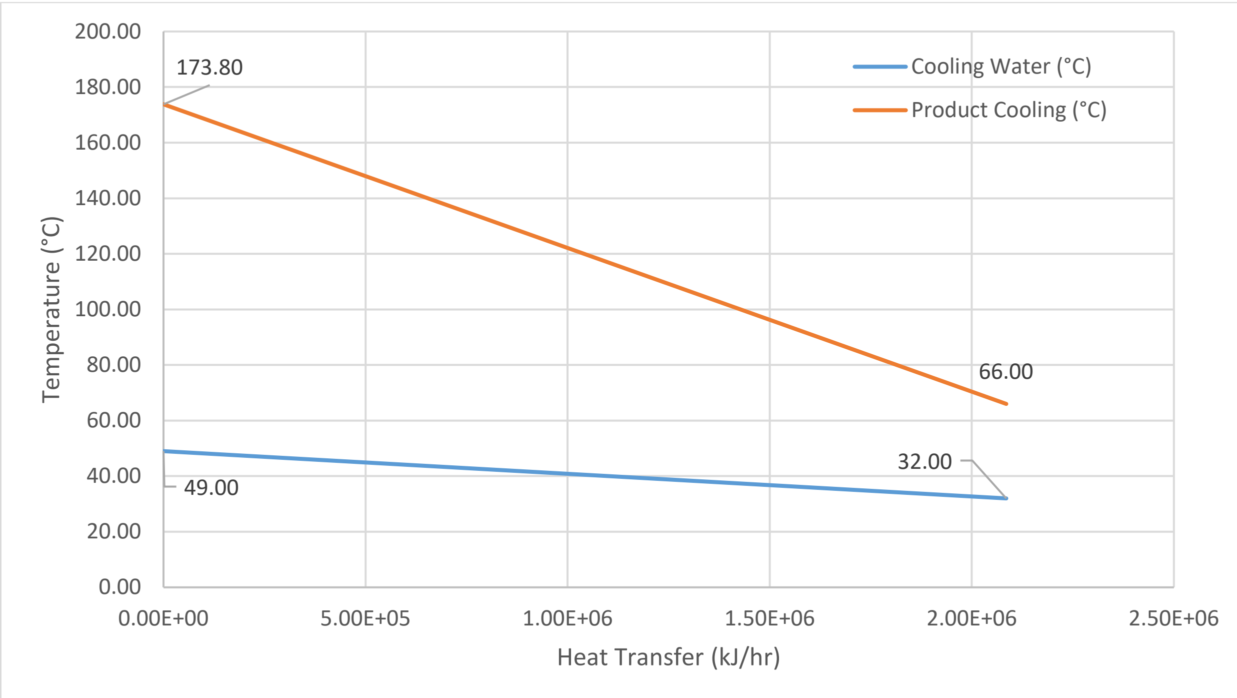


Figure 24: T-Q Diagram for E-203

Table 64: HEX Calculations for E-203

Q (kJ/h)	Cooling Water (°C)	Product Cooling (°C)	Delta T
0.00E+00	49.00	173.80	124.8
2.09E+06	32.00	66.00	34
2.09E+06		LMTD	69.82725
Function	Cooling Water	Cooling Wastewater	
HEX Side	Tube Side	Shell Side	

Variable	Value	Units
Q	579.23	kW
Uo=	857.14	W/(m^2)*K
F	0.9751	2 shell pass, 4 tube pass
LMTD	69.8	°C

Variable	Value	Units
hi	5000	W/(m^2)*K
Rfi	0.0002	((m^2)*K)/W
ho	1500	W/(m^2)*K
Rfo	0.0001	((m^2)*K)/W

A=	9.92	m^2
A=	106.83	ft^2
A*1.1=	10.92	m^2
A*1.1=	117.51	ft^2

Table 65: E-203 Cost Calculations

Heat Exchanger Costing Calculator						
Variable	Equation	U-Tube	Floating Head	Fixed Tube	Bayonet	Units
C_{po} Pg. 910	$\log_{10}(C_P^0)=K_1+K_2\log_{10}(A)+K3\log_{10}(A)]^2$	$ 13,845.07	$ 19,531.01	$ 15,354.82	$ 23,967.94	$
K_1	Table A.1 in Turton	4.1884	4.8306	4.3247	4.2768	N/A
K_2	Table A.1 in Turton	-0.2503	-0.8509	-0.303	-0.0495	N/A
K_3	Table A.1 in Turton	0.1974	0.3187	0.1634	0.1431	N/A
Finding F_{BM}	FBM=B1+B2*FM*FP	3.57	3.57	3.57	3.57	N/A
Table A.4 pg. 934: B1	B1	1.63	1.63	1.63	1.63	N/A
Table A.4 pg. 935: B2	B2	1.66	1.66	1.66	1.66	N/A
Pg. 931 F_M	Carbon Steel shell and tube	1	1	1	1	N/A
Fp (A.3)	$\log_{10}$(FP)=C1+C2log10(P)+ (C3((log10(P))^2)	1.17	1.17	1.17	1.17	N/A
Pressure	Converting psia to barg	40.00	40.00	40.00	40.00	barg
C1		0.03881	0.03881	0.03881	0.03881	$
C2		-0.11272	-0.11272	-0.11272	-0.11272	N/A
C3		0.08183	0.08183	0.08183	0.08183	N/A
$C_{BM\ 2001}$	$C_{BM}=Cp^o*F_{BM}$	$ 49,461.34	$ 69,774.31	$ 54,854.91	$ 85,625.19	$
CAPCOST	CEPCI 2001: 397					N/A
	CEPCI 2017: 566.6					N/A
$C_{BM\ 2017}$	CBM2017=CBM2001*(566.6/397)	$ 70,591.42	$ 99,582.17	$ 78,289.15	$122,204.61	$
Contingency	CBM2017*0.15	$ 10,588.71	$ 14,937.33	$ 11,743.37	$ 18,330.69	$
Fees	CBM2017*0.03	$ 2,117.74	$ 2,987.47	$ 2,348.67	$ 3,666.14	$
C_{TM}		$ 83,297.88	$117,506.96	$ 92,381.19	$144,201.44	$
Operating Cost:						
Heat Transferred		579.23	kW			
CW energy required		1.74E+10	kJ/yr			
Water flow rate		0.452070507	kmol/s			
Volumetric flow rate		0.01	cubic meters/s	For both HEX	Rounded	
Water Required per year		245,455.17	cubic meters/yr	490,910.34	491,000.00	
Annual Cost of CW		$ 12,272.76	$/yr	24,545.52	25,000.00	
PWC of CW @ 8% ROR and 20 yr lifetime		$ 120,493.94		240,987.88	241,000.00	
Total PWC of HEX		$ 212,875.14		425,750.27	426,000.00	

Appendix C: Tower Calculations

T-201

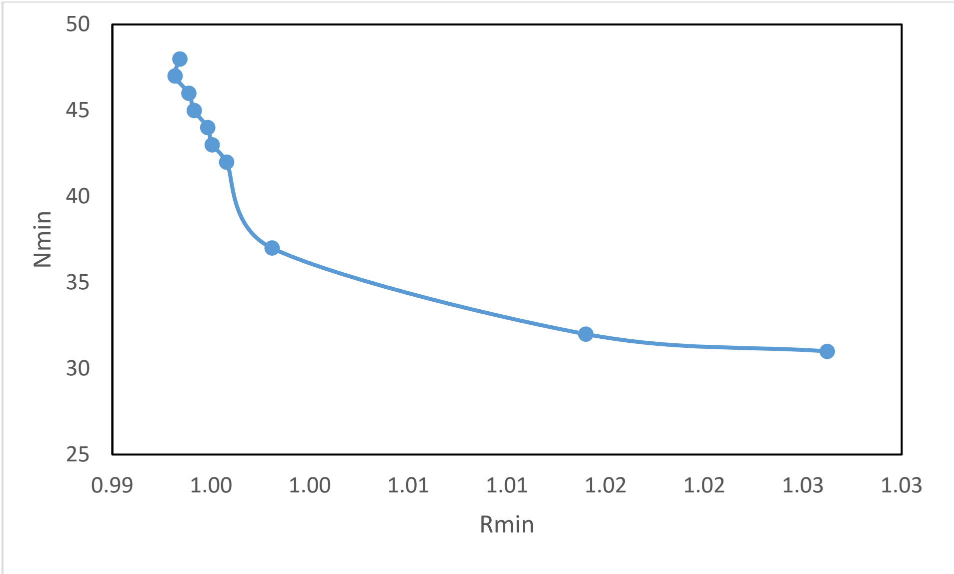


Figure 25: A plot of Nmin versus Rmin showing an optimum number of actual trays in the 40-46 tray region.

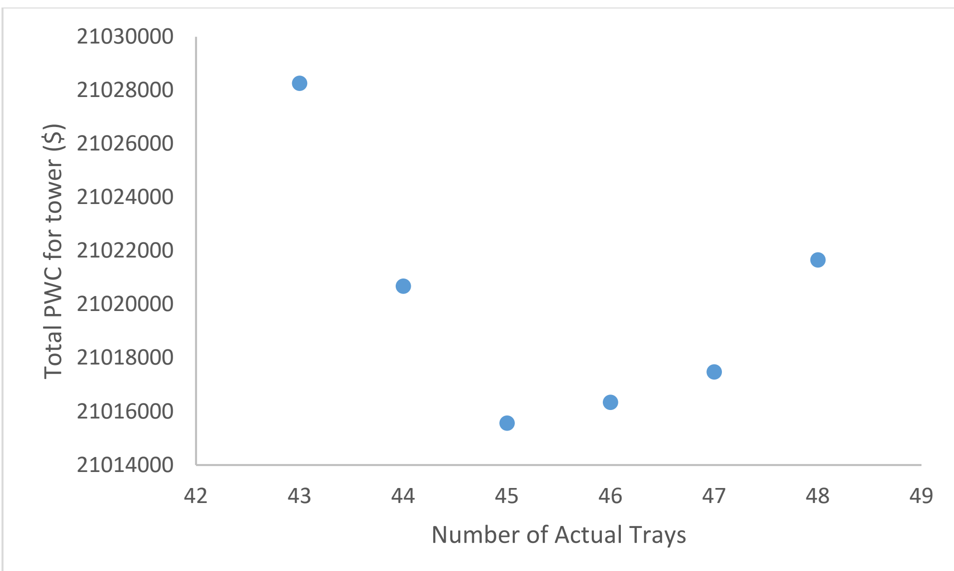


Figure 26: Cost of distillation columns at varying number of stages. The cheapest Tower is one that has 23 actual stages.

Table 66: Reflux Ratio at various Feed stages for the 32 stage design

Number of actual stages	Feed Stage	Reflux Ratio
45	27	0.9946
	28	0.9943
	29	0.9942
	30	0.9942
	31	0.9943
	32	0.9943
	33	0.9972
	34	1.0005

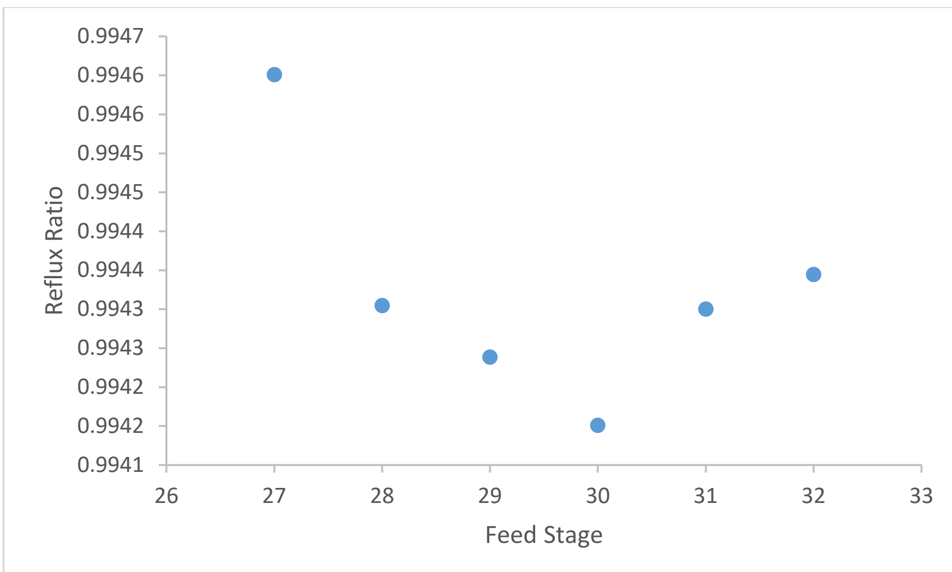


Figure 27: A plot of Reflux Ratio vs. Feed Stage for the 32 actual tray simulation. The optimal Feed stage is stage 21.

Table 67 Tower properties

Select Carbon Steel			
Parameter	Equation	Value	Units
Volume: Area*Height(ft^3)	Volume: Area*Height (ft^3)	515.75	ft^3
Height	Column Height= $h_{top}+h_{trays}+h_{btms}$	73.00	Ft
h_{top}	Tray Spacing plus 1 ft	2	Ft
h_{trays}	From Hysys	64	Ft
h_{btms}	(3 ft for vapor/liquid disengagement and 4 ft for holdup)	7	Ft
Area	Area= (pi/4)* (Diameter)^2	7.07	ft^2
Diameter	From Hysys	3	Ft
Pressure	From Hysys	12.35	Barg
Number of theoretical stages	From Hysys	23	Stages
Number of actual trays	From Hysys	32	Trays
Efficiency	From Hysys	72	%

Table 68: Tray Parameters

Sieve Trays			
Parameter	Source	Value	Units
Area, A	Area (m^2) from HYSYS	7.065	ft^2
Diameter	From HYSYS	3	Ft

Table 69:Condenser E-201 A/B

Parameter	Value	Units
Area	494.8	ft^2
Area with safety factor	544.3	ft^2
LMTD	99	N/A
U	850	W/m^2K
Duty	$1.35*10^7$	kJ/hr
Pressure	12.1	Barg
Temperature	131.6	C

Table 70: Reboiler E-202 A/B Properties

Parameter	value	units
Duty	$1.41*10^7$	(KJ/hr)
Temperature	173.8	C
Steam Pressure	41	barg
Steam Temperature	254	C
Latent Heat	2019	KJ/Kg
LMTD	80.2	N/A
U	850	(W/m^2K)
F	0.9751	N/A
A	633.9	ft^2
A with Safety Factor	697.3	ft^2
Pressure	14.3	barg

Appendix D: Pump Sizing Calculations

Table 71: Calculations for P-102

P - 102 A/B/C/D		
Parameter	**Value**	**Units**
Flow Rate:	147.0	gal/min
Head:	64.0	ft
Pump Selection	Single-Stage Centrifugal	
Pressure	169.5	psia
Hydraulic hp	1.11	kW
Efficiency	50	%
Break hp	2.22	kW

Table 72: Calculations for P-103

P - 103 A/B/C/D		
Parameter	**Value**	**Units**
Flow Rate:	71.3	gal/min
Head:	26.25	ft
Pump Selection	Single-Stage Centrifugal	
Pressure	164.8	psia
Hydraulic hp	0.29	kW
Efficiency	35	%
Break hp	0.82	kW

Table 73: Calculations for P-201

P - 201 A/B/C/D		
What we are Finding	**Value**	**Units**
Flow Rate:	94.6	gal/min
Head:	88.75	ft
Pump Selection	Single-Stage Centrifugal	
Pressure	150.3	psia
Hydraulic hp	1.16	kW
Efficiency	50	%
Break hp	2.32	kW

Table 74: Calculations for P-301

P - 301 A/B		
Parameter	**Value**	**Units**
Flow Rate:	**0.11**	**gal/min**
Head:	**45**	**ft**
Pump Selection	**Positive Displacement**	
Pressure	**75**	**psia**
Hydraulic hp	**0.003**	**kW**
Efficiency	**35**	**%**
Break hp	**0.009**	**kW**

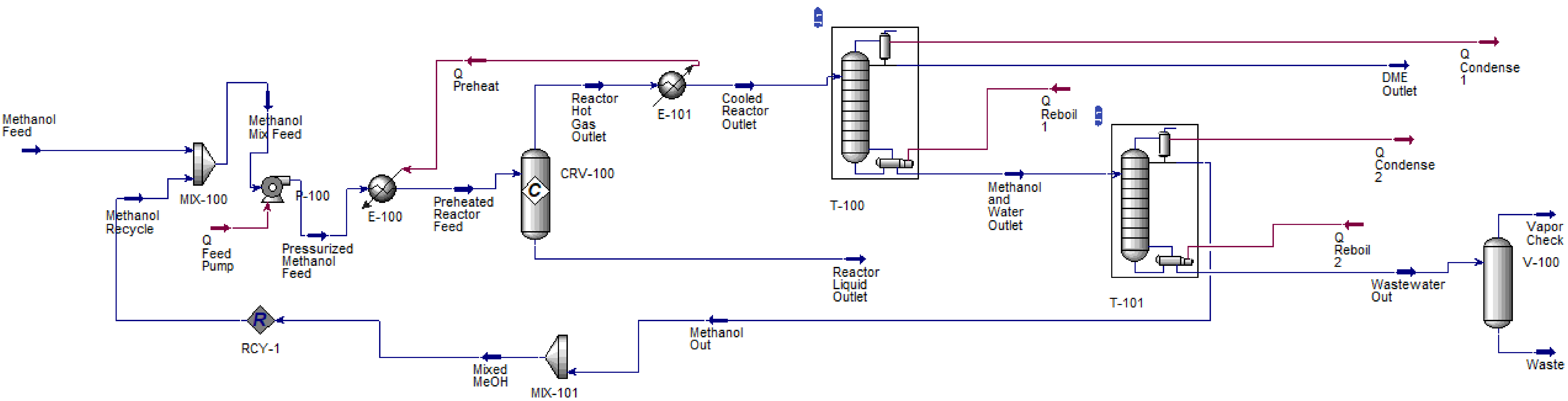
Appendix E: Cash Flow Table

Table 75: Complete Cash Flow Table

Project Title: DME Synthesis for Transportation	*Propylene Splitter*					Fixed Cap	$ 54,313,228	Operating Labor	$ 844,480	
Corporate financial situation	*Expense*					Cost of MeOH/gal	$2.26			
Corporate Rate of Return i*	8%					Operating Cost	$ 6,232,445			
Other Relevant Project Info	*10 Year MACRS Depreciation*					Working Capital	$ 15,646,984			
End of Year	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027
Production DME (U.S Gallons/year)						91,250,000	91,250,000	91,250,000	91,250,000	91,250,000
x Sales Price ($/U.S Gallon)						$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26
+Net Revenue						$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000
- Raw Material Costs (Lubricant+ Methanol)						$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)
Methanol (U.S Gallons/year)						94,224,278	94,224,278	94,224,278	94,224,278	94,224,278
x cost Price ($/U.S Gallons)						1.49	1.49	1.49	1.49	1.49
x cost Shipment cost ($/U.S Gallon)						0.02	0.02	0.02	0.02	0.02
Lubrication (lb/year)						410,143.20	410,143.20	410,143.20	410,143.20	410,143.20
Lubrication Cost ($/lb)						1.65	1.65	1.65	1.65	1.65
Lubrication (U.S Gallons/year)						70208	70208	70208	70208	70208
Lubrication Shipping Cost ($/U.S Gallon)						0.02	0.02	0.02	0.02	0.02
Total Lubrication Cost $/year						678,140.44	678,140.44	678,140.44	678,140.44	678,140.44
Waste Water Treament (U.S Gallons/year)						29,807,652	29,807,652	29,807,652	29,807,652	29,807,652
- Waste Water Treament ($/year)						$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)
- Other Direct Costs						$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)
- Fixed Costs						$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)
- General Expenses						$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)
- Utilities						$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)
- Manual Labor Costs						$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)
- Depreciation						$ (5,431,323)	$ (9,776,381)	$ (7,821,105)	$ (6,256,884)	$ (5,007,680)
Depreciation Ratio, 10-year MACRS						0.1000	0.1800	0.1440	0.1152	0.0922
- Loss Forward						$ -	$ -	$ -	$ -	$ -
- Write Off										
Taxable Income						$ 5,590,874	$ 1,245,816	$ 3,201,092	$ 4,765,313	$ 6,014,517
- Tax @ 40%						$ (2,236,350)	$ (498,326)	$ (1,280,437)	$ (1,906,125)	$ (2,405,807)
Net Income						$ 3,354,525	$ 747,490	$ 1,920,655	$ 2,859,188	$ 3,608,710
+ Depreciation						$ 5,431,323	$ 9,776,381	$ 7,821,105	$ 6,256,884	$ 5,007,680
+ Loss Forward						$ -	$ -	$ -	$ -	$ -
+ Write Off						$ -	$ -	$ -	$ -	$ -
- Working Capital					$ (15,646,984)	$ -	$ -	$ -	$ -	$ -
- Fixed Capital equipment cost	$ (3,258,794)	$ (14,121,439)	$ (10,862,646)	$ (15,207,704)	$ (10,862,646)	$ -	$ -	$ -	$ -	$ -
Fixed Capital expense schedule	0.06	0.26	0.20	0.28	0.20	0	0	0	0	0
Cash Flow	$ (3,258,794)	$ (14,121,439)	$ (10,862,645)	$ (15,207,703)	$ (26,509,629)	$ 8,785,847	$ 10,523,871	$ 9,741,760	$ 9,116,072	$ 8,616,390
Discount Factor(P/F_i',n)	1.0000	0.9259	0.8573	0.7938	0.7350	0.6806	0.6302	0.5835	0.5403	0.5002
Discounted Cash Flow	$ (3,258,794)	$ (13,075,406)	$ (9,312,968)	$ (12,072,365)	$ (19,485,369)	$ 5,979,500	$ 6,631,824	$ 5,684,223	$ 4,925,130	$ 4,310,340
NPV @i' =	$ 2,488,958	$ 2,500,000								
DCFROR =	8.52%	8.5%								

End of Year	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042
Production DME (U.S Gallons/year)	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000	91,250,000
x Sales Price ($/U.S Gallon)	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26	$ 2.26
+Net Revenue	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000	$ 206,225,000
- Raw Material Costs (Lubricant+ Methanol)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)	$ (141,600,519)
Methanol (U.S Gallons/year)	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278	94,224,278
x cost Price ($/U.S Gallons)	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49
x cost Shipment cost ($/U.S Gallon)	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Lubrication (lb/year)	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20	410,143.20
Lubrication Cost ($/lb)	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65	1.65
Lubrication (U.S Gallons/year)	70208	70208	70208	70208	70208	70208	70208	70208	70208	70208	70208	70208	70208	70208	70208
Lubrication Shipping Cost ($/U.S Gallon)	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Total Lubrication Cost $/year	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44	678,140.44
Waste Water Treament (U.S Gallons/year)	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652	29,807,652
- Waste Water Treament ($/year)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)	$ (131,750)
- Other Direct Costs	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)	$ (9,937,528)
- Fixed Costs	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)	$ (4,291,191)
- General Expenses	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)	$ (32,164,889)
- Utilities	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)	$ (6,232,445)
- Manual Labor Costs	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)	$ (844,480)
- Depreciation	$ (4,002,885)	$ (3,557,516)	$ (3,557,516)	$ (3,562,948)	$ (3,557,516)	$ (1,781,474)	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Depreciation Ratio, 10-year MACRS	0.0737	0.0655	0.0655	0.0656	0.0655	0.0328									
- Loss Forward	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
- Write Off															$ (15,646,984)
Taxable Income	$ 7,019,312	$ 7,464,681	$ 7,464,681	$ 7,459,249	$ 7,464,681	$ 9,240,723	$ 11,022,197	$ 11,022,197	$ 11,022,197	$ 11,022,197	$ 11,022,197	$ 11,022,197	$ 11,022,197	$ 11,022,197	$ (4,624,787)
- Tax @ 40%	$ (2,807,725)	$ (2,985,872)	$ (2,985,872)	$ (2,983,700)	$ (2,985,872)	$ (3,696,289)	$ (4,408,879)	$ (4,408,879)	$ (4,408,879)	$ (4,408,879)	$ (4,408,879)	$ (4,408,879)	$ (4,408,879)	$ (4,408,879)	$ -
Net Income	$ 4,211,587	$ 4,478,808	$ 4,478,808	$ 4,475,550	$ 4,478,808	$ 5,544,434	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ (4,624,787)
+ Depreciation	$ 4,002,885	$ 3,557,516	$ 3,557,516	$ 3,562,948	$ 3,557,516	$ 1,781,474	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
+ Loss Forward	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
+ Write Off	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ 15,646,984
- Working Capital	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
- Fixed Capital equipment cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Fixed Capital expense schedule	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cash Flow	$ 8,214,472	$ 8,036,325	$ 8,036,325	$ 8,038,497	$ 8,036,325	$ 7,325,908	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 6,613,318	$ 11,022,197
Discount Factor(P/F_i',n)	0.4632	0.4289	0.3971	0.3677	0.3405	0.3152	0.2919	0.2703	0.2502	0.2317	0.2145	0.1987	0.1839	0.1703	0.1577
Discounted Cash Flow	$ 3,804,890	$ 3,446,642	$ 3,191,335	$ 2,955,739	$ 2,736,055	$ 2,309,432	$ 1,930,365	$ 1,787,375	$ 1,654,976	$ 1,532,386	$ 1,418,876	$ 1,313,774	$ 1,216,457	$ 1,126,349	$ 1,738,193

Appendix F: HYSYS Simulation Flow Sheet



Supplementary Information: HYSYS Report

OKLAHOMA STATE UNIVERSIT Case Name: Half Flow Optimization.hsc

Bedford, MA

USA Unit Set: Field

Date/Time: Tue Mar 13 09:25:57 2018

All Streams (Case (Main)): Worksheet, Attachments

Material Stream: Methanol Feed Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	77.00	77.00
Pressure: (psia)	14.65	14.65
Molar Flow (lbmole/hr)	1160	1160
Mass Flow (lb/hr)	3.715e+004	3.715e+004
Std Ideal Liq Vol Flow (barrel/day)	3196	3196
Molar Enthalpy (Btu/lbmole)	-1.030e+005	-1.030e+005
Molar Entropy (Btu/lbmole-F)	11.15	11.15
Heat Flow (Btu/hr)	-1.195e+008	-1.195e+008
Liq Vol Flow @Std Cond (barrel/day)	3193	3193

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	32.02	32.02
Molar Density (lbmole/ft3)	1.533	1.533
Mass Density (lb/ft3)	49.08	49.08
Act. Volume Flow (barrel/day)	3235	3235
Mass Enthalpy (Btu/lb)	-3216	-3216
Mass Entropy (Btu/lb-F)	0.3484	0.3484
Heat Capacity (Btu/lbmole-F)	27.57	27.57
Mass Heat Capacity (Btu/lb-F)	0.8610	0.8610
LHV Molar Basis (Std) (Btu/lbmole)	2.738e+005	2.738e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.091e+005	3.091e+005
HHV Mass Basis (Std) (Btu/lb)	9654	9654
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8553	8553
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	0.1067	---

Partial Pressure of CO2 (psia) 0.0000 ---

Cost Based on Flow (Cost/s) 0.0000 0.0000

Act. Gas Flow (ACFM) --- ---

Avg. Liq. Density (lbmole/ft3) 1.552 1.552

Specific Heat (Btu/lbmole-F) 27.57 27.57

Std. Gas Flow (MMSCFD) 10.55 10.55

Std. Ideal Liq. Mass Density (lb/ft3) 49.69 49.69

Act. Liq. Flow (USGPS) 1.573 1.573

Z Factor 1.659e-003 1.659e-003

Watson K 10.63 10.63

User Property --- ---

Partial Pressure of H2S (psia) 0.0000 ---

Cp/(Cp - R) 1.078 1.078

Cp/Cv 1.373 1.373

Heat of Vap. (Btu/lbmole) 1.518e+004 ---

Kinematic Viscosity (cSt) 0.6933 0.6933

Liq. Mass Density (Std. Cond) (lb/ft3) 49.73 49.73

Liq. Vol. Flow (Std. Cond) (barrel/day) 3193 3193

Liquid Fraction 1.000 1.000

Molar Volume (ft3/lbmole) 0.6523 0.6523

Mass Heat of Vap. (Btu/lb) 474.1 ---

Phase Fraction [Molar Basis] 0.0000 1.0000

Surface Tension (dyne/cm) 29.67 29.67

Thermal Conductivity (Btu/hr-ft-F) 0.1041 0.1041

Viscosity (cP) 0.5451 0.5451

Cv (Semi-Ideal) (Btu/lbmole-F) 25.58 25.58

Mass Cv (Semi-Ideal) (Btu/lb-F) 0.7990 0.7990

Cv (Btu/lbmole-F) 20.08 20.08

Mass Cv (Btu/lb-F) 0.6272 0.6272

Cv (Ent. Method) (Btu/lbmole-F) --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- ---

Cp/Cv (Ent. Method) --- ---

Reid VP at 37.8 C (psia) 4.641 4.641

True VP at 37.8 C (psia) 4.636 4.636

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 3193 3193

Viscosity Index -2.307 ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	2.0621	0.0018	37.1487	0.0010	2.5488	0.0008
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	1158.2189	0.9982	37111.5347	0.9990	3193.5001	0.9992
Total	1160.2810	1.0000	37148.6834	1.0000	3196.0489	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	2.0621	0.0018	37.1487	0.0010	2.5488	0.0008
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	1158.2189	0.9982	37111.5347	0.9990	3193.5001	0.9992
Total	1160.2810	1.0000	37148.6834	1.0000	3196.0489	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	---	---	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Mixer: MIX-100		

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Preheated Reactor Feed Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature: (F)	536.0	536.0
Pressure: (psia)	500.0	500.0
Molar Flow (lbmole/hr)	1668	1668
Mass Flow (lb/hr)	5.299e+004	5.299e+004
Std Ideal Liq Vol Flow (barrel/day)	4550	4550
Molar Enthalpy (Btu/lbmole)	-8.101e+004	-8.101e+004
Molar Entropy (Btu/lbmole-F)	42.56	42.56
Heat Flow (Btu/hr)	-1.351e+008	-1.351e+008
Liq Vol Flow @Std Cond (barrel/day)	4546	4546

PROPERTIES

	Overall	Vapour Phase
Molecular Weight	31.77	31.77
Molar Density (lbmole/ft3)	4.679e-002	4.679e-002
Mass Density (lb/ft3)	1.487	1.487
Act. Volume Flow (barrel/day)	1.524e+005	1.524e+005
Mass Enthalpy (Btu/lb)	-2550	-2550

Mass Entropy (Btu/lb-F)	1.340	1.340
Heat Capacity (Btu/lbmole-F)	15.04	15.04
Mass Heat Capacity (Btu/lb-F)	0.4733	0.4733
LHV Molar Basis (Std) (Btu/lbmole)	2.690e+005	2.690e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.040e+005	3.040e+005
HHV Mass Basis (Std) (Btu/lb)	9567	9567
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8468	8468
Phase Fraction [Vol. Basis]	1.000	1.000
Phase Fraction [Mass Basis]	1.000	1.000
Phase Fraction [Act. Vol. Basis]	1.000	1.000
Mass Exergy (Btu/lb)	171.6	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	594.1	594.1
Avg. Liq. Density (lbmole/ft3)	1.567	1.567
Specific Heat (Btu/lbmole-F)	15.04	15.04
Std. Gas Flow (MMSCFD)	15.16	15.16
Std. Ideal Liq. Mass Density (lb/ft3)	49.78	49.78
Act. Liq. Flow (USGPS)	---	---
Z Factor	1.000	1.000
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.152	1.152
Cp/Cv	1.152	1.152
Heat of Vap. (Btu/lbmole)	1.050e+004	---
Kinematic Viscosity (cSt)	0.5549	0.5549
Liq. Mass Density (Std. Cond) (lb/ft3)	49.83	49.83
Liq. Vol. Flow (Std. Cond) (barrel/day)	4546	4546
Liquid Fraction	0.0000	0.0000
Molar Volume (ft3/lbmole)	21.37	21.37
Mass Heat of Vap. (Btu/lb)	330.6	---
Phase Fraction [Molar Basis]	1.0000	1.0000
Surface Tension (dyne/cm)	---	---
Thermal Conductivity (Btu/hr-ft-F)	2.205e-002	2.205e-002
Viscosity (cP)	1.322e-002	1.322e-002
Cv (Semi-Ideal) (Btu/lbmole-F)	13.05	13.05
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.4108	0.4108
Cv (Btu/lbmole-F)	13.05	13.05
Mass Cv (Btu/lb-F)	0.4108	0.4108
Cv (Ent. Method) (Btu/lbmole-F)	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---
Cp/Cv (Ent. Method)	---	---
Reid VP at 37.8 C (psia)	4.669	4.669
True VP at 37.8 C (psia)	4.615	4.615

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 4546 4546

Viscosity Index --- ---

COMPOSITION

Overall Phase Vapour Fraction 1.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	32.6646	0.0196	588.4568	0.0111	40.3746	0.0089
diM-Ether	0.4174	0.0003	19.2302	0.0004	1.9644	0.0004
Methanol	1634.9186	0.9802	52385.8982	0.9885	4507.8807	0.9907
Total	1668.0006	1.0000	52993.5852	1.0000	4550.2197	1.0000

Vapour Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	32.6646	0.0196	588.4568	0.0111	40.3746	0.0089
diM-Ether	0.4174	0.0003	19.2302	0.0004	1.9644	0.0004
Methanol	1634.9186	0.9802	52385.8982	0.9885	4507.8807	0.9907
Total	1668.0006	1.0000	52993.5852	1.0000	4550.2197	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	---	---	---
diM-Ether	---	---	---
Methanol	---	---	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Conversion Reactor: CRV-100	Heater: E-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Reactor Hot Gas Outlet Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Vapour Phase	Aqueous Phase
Vapour / Phase Fraction	1.0000	1.0000	0.0000
Temperature: (F)	717.6	717.6	717.6
Pressure: (psia)	500.0	500.0	500.0
Molar Flow (lbmole/hr)	1668	1668	0.0000

Mass Flow (lb/hr)	5.299e+004	5.299e+004	0.0000
Std Ideal Liq Vol Flow (barrel/day)	4797	4797	0.0000
Molar Enthalpy (Btu/lbmole)	-8.101e+004	-8.101e+004	-8.970e+004
Molar Entropy (Btu/lbmole-F)	45.95	45.95	43.68
Heat Flow (Btu/hr)	-1.351e+008	-1.351e+008	0.0000
Liq Vol Flow @Std Cond (barrel/day)	4810	4810	0.0000

PROPERTIES

	Overall	Vapour Phase	Aqueous Phase
Molecular Weight	31.77	31.77	26.70
Molar Density (lbmole/ft3)	3.958e-002	3.958e-002	3.468
Mass Density (lb/ft3)	1.257	1.257	92.59
Act. Volume Flow (barrel/day)	1.802e+005	1.802e+005	0.0000
Mass Enthalpy (Btu/lb)	-2550	-2550	-3360
Mass Entropy (Btu/lb-F)	1.446	1.446	1.636
Heat Capacity (Btu/lbmole-F)	17.38	17.38	16.75
Mass Heat Capacity (Btu/lb-F)	0.5471	0.5471	0.6274
LHV Molar Basis (Std) (Btu/lbmole)	2.768e+005	2.768e+005	1.745e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.117e+005	3.117e+005	2.031e+005
HHV Mass Basis (Std) (Btu/lb)	9812	9812	7607
CO2 Loading	---	---	---
CO2 App ML Con (lbmole/ft3)	---	---	---
CO2 App WT Con (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	8713	8713	6538
Phase Fraction [Vol. Basis]	1.000	1.000	---
Phase Fraction [Mass Basis]	1.000	1.000	0.0000
Phase Fraction [Act. Vol. Basis]	1.000	1.000	0.0000
Mass Exergy (Btu/lb)	244.8	---	---
Partial Pressure of CO2 (psia)	0.0000	---	---
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	702.4	702.4	---
Avg. Liq. Density (lbmole/ft3)	1.486	1.486	1.885
Specific Heat (Btu/lbmole-F)	17.38	17.38	16.75
Std. Gas Flow (MMSCFD)	15.16	15.16	0.0000
Std. Ideal Liq. Mass Density (lb/ft3)	47.22	47.22	50.31
Act. Liq. Flow (USGPS)	0.0000	---	0.0000
Z Factor	---	1.000	1.141e-002
Watson K	11.19	11.19	11.18
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	---	---
Cp/(Cp - R)	1.129	1.129	1.135
Cp/Cv	1.129	1.129	1.135
Heat of Vap. (Btu/lbmole)	1.256e+004	---	---
Kinematic Viscosity (cSt)	0.9393	0.9393	9.262e-002
Liq. Mass Density (Std. Cond) (lb/ft3)	47.10	47.10	51.09
Liq. Vol. Flow (Std. Cond) (barrel/day)	4810	4810	0.0000
Liquid Fraction	0.0000	0.0000	1.000
Molar Volume (ft3/lbmole)	25.27	25.27	0.2883

Mass Heat of Vap. (Btu/lb) 395.5 --- ---

Phase Fraction [Molar Basis] 1.0000 1.0000 0.0000

Surface Tension (dyne/cm) --- --- 0.0000

Thermal Conductivity (Btu/hr-ft-F) 3.059e-002 3.059e-002 7.587e-002

Viscosity (cP) 1.892e-002 1.892e-002 0.1374

Cv (Semi-Ideal) (Btu/lbmole-F) 15.40 15.40 14.76

Mass Cv (Semi-Ideal) (Btu/lb-F) 0.4846 0.4846 0.5530

Cv (Btu/lbmole-F) 15.40 15.40 14.76

Mass Cv (Btu/lb-F) 0.4846 0.4846 0.5530

Cv (Ent. Method) (Btu/lbmole-F) --- --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- --- ---

Cp/Cv (Ent. Method) --- --- ---

Reid VP at 37.8 C (psia) 65.93 65.93 62.96

True VP at 37.8 C (psia) 40.34 40.34 24.81

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 4810 4810 0.0000

Viscosity Index --- --- ---

COMPOSITION

Overall Phase Vapour Fraction 1.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	610.1179	0.3658	10991.3349	0.2074	754.1266	0.1572
diM-Ether	577.8707	0.3464	26621.9235	0.5024	2719.4883	0.5669
Methanol	480.0121	0.2878	15380.4997	0.2902	1323.5138	0.2759
Total	1668.0006	1.0000	52993.7581	1.0000	4797.1286	1.0000

Vapour Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	610.1179	0.3658	10991.3349	0.2074	754.1266	0.1572
diM-Ether	577.8707	0.3464	26621.9235	0.5024	2719.4883	0.5669
Methanol	480.0121	0.2878	15380.4997	0.2902	1323.5138	0.2759
Total	1668.0006	1.0000	52993.7581	1.0000	4797.1286	1.0000

Aqueous Phase Phase Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0000	0.5929	0.0000	0.4001	0.0000	0.3231
diM-Ether	0.0000	0.2118	0.0000	0.3656	0.0000	0.4395
Methanol	0.0000	0.1952	0.0000	0.2343	0.0000	0.2373
Total	0.0000	1.0000	0.0000	1.0000	0.0000	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.6169	---	0.6169
diM-Ether	1.635	---	1.635
Methanol	1.474	---	1.474

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Cooler: E-101	Conversion Reactor: CRV-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Reactor Liquid Outlet Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Vapour Phase	Aqueous Phase
Vapour / Phase Fraction	0.0000	0.0000	1.0000
Temperature: (F)	717.6	717.6	717.6
Pressure: (psia)	500.0	500.0	500.0
Molar Flow (lbmole/hr)	0.0000	0.0000	0.0000
Mass Flow (lb/hr)	0.0000	0.0000	0.0000
Std Ideal Liq Vol Flow (barrel/day)	0.0000	0.0000	0.0000
Molar Enthalpy (Btu/lbmole)	-8.970e+004	-8.101e+004	-8.970e+004
Molar Entropy (Btu/lbmole-F)	43.68	45.95	43.68
Heat Flow (Btu/hr)	0.0000	0.0000	0.0000
Liq Vol Flow @Std Cond (barrel/day)	0.0000	0.0000	0.0000

PROPERTIES

	Overall	Vapour Phase	Aqueous Phase
Molecular Weight	26.70	31.77	26.70
Molar Density (lbmole/ft3)	3.468	3.958e-002	3.468
Mass Density (lb/ft3)	92.59	1.257	92.59
Act. Volume Flow (barrel/day)	0.0000	0.0000	0.0000
Mass Enthalpy (Btu/lb)	-3360	-2550	-3360
Mass Entropy (Btu/lb-F)	1.636	1.446	1.636
Heat Capacity (Btu/lbmole-F)	16.75	17.38	16.75
Mass Heat Capacity (Btu/lb-F)	0.6274	0.5471	0.6274
LHV Molar Basis (Std) (Btu/lbmole)	1.745e+005	2.768e+005	1.745e+005
HHV Molar Basis (Std) (Btu/lbmole)	2.031e+005	3.117e+005	2.031e+005
HHV Mass Basis (Std) (Btu/lb)	7607	9812	7607
CO2 Loading	---	---	---
CO2 App ML Con (lbmole/ft3)	---	---	---
CO2 App WT Con (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	6538	8713	6538
Phase Fraction [Vol. Basis]	---	---	1.000
Phase Fraction [Mass Basis]	0.0000	0.0000	1.000

Phase Fraction [Act. Vol. Basis] --- --- ---

Mass Exergy (Btu/lb) 159.4 --- ---

Partial Pressure of CO2 (psia) 0.0000 --- ---

Cost Based on Flow (Cost/s) 0.0000 0.0000 0.0000

Act. Gas Flow (ACFM) --- --- ---

Avg. Liq. Density (lbmole/ft3) --- --- ---

Specific Heat (Btu/lbmole-F) 16.75 17.38 16.75

Std. Gas Flow (MMSCFD) 0.0000 0.0000 0.0000

Std. Ideal Liq. Mass Density (lb/ft3) 50.31 47.22 50.31

Act. Liq. Flow (USGPS) 0.0000 --- 0.0000

Z Factor --- 1.000 1.141e-002

Watson K 11.18 11.19 11.18

User Property --- --- ---

Partial Pressure of H2S (psia) 0.0000 --- ---

Cp/(Cp - R) 1.135 1.129 1.135

Cp/Cv 1.135 1.129 1.135

Heat of Vap. (Btu/lbmole) 1.342e+004 --- ---

Kinematic Viscosity (cSt) 9.262e-002 0.9393 9.262e-002

Liq. Mass Density (Std. Cond) (lb/ft3) 51.09 47.10 51.09

Liq. Vol. Flow (Std. Cond) (barrel/day) 0.0000 0.0000 0.0000

Liquid Fraction 1.000 0.0000 1.000

Molar Volume (ft3/lbmole) 0.2883 25.27 0.2883

Mass Heat of Vap. (Btu/lb) 502.8 --- ---

Phase Fraction [Molar Basis] 0.0000 0.0000 1.0000

Surface Tension (dyne/cm) --- --- ---

Thermal Conductivity (Btu/hr-ft-F) 7.587e-002 3.059e-002 7.587e-002

Viscosity (cP) 0.1374 1.892e-002 0.1374

Cv (Semi-Ideal) (Btu/lbmole-F) 14.76 15.40 14.76

Mass Cv (Semi-Ideal) (Btu/lb-F) 0.5530 0.4846 0.5530

Cv (Btu/lbmole-F) 14.76 15.40 14.76

Mass Cv (Btu/lb-F) 0.5530 0.4846 0.5530

Cv (Ent. Method) (Btu/lbmole-F) --- --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- --- ---

Cp/Cv (Ent. Method) --- --- ---

Reid VP at 37.8 C (psia) 62.96 65.93 62.96

True VP at 37.8 C (psia) 24.81 40.34 24.81

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 0.0000 0.0000 0.0000

Viscosity Index --- --- ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0000	0.5929	0.0000	0.4001	0.0000	0.3231
diM-Ether	0.0000	0.2118	0.0000	0.3656	0.0000	0.4395
Methanol	0.0000	0.1952	0.0000	0.2343	0.0000	0.2373
Total	0.0000	1.0000	0.0000	1.0000	0.0000	1.0000

Vapour Phase Phase Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0000	0.3658	0.0000	0.2074	0.0000	0.1572
diM-Ether	0.0000	0.3464	0.0000	0.5024	0.0000	0.5669
Methanol	0.0000	0.2878	0.0000	0.2902	0.0000	0.2759
Total	0.0000	1.0000	0.0000	1.0000	0.0000	1.0000

Aqueous Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0000	0.5929	0.0000	0.4001	0.0000	0.3231
diM-Ether	0.0000	0.2118	0.0000	0.3656	0.0000	0.4395
Methanol	0.0000	0.1952	0.0000	0.2343	0.0000	0.2373
Total	0.0000	1.0000	0.0000	1.0000	0.0000	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.6169	---	0.6169
diM-Ether	1.635	---	1.635
Methanol	1.474	---	1.474

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Conversion Reactor: CRV-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Cooled Reactor Outlet Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	202.0	202.0
Pressure: (psia)	500.0	500.0
Molar Flow (lbmole/hr)	1668	1668
Mass Flow (lb/hr)	5.299e+004	5.299e+004
Std Ideal Liq Vol Flow (barrel/day)	4797	4797
Molar Enthalpy (Btu/lbmole)	-1.016e+005	-1.016e+005
Molar Entropy (Btu/lbmole-F)	28.90	28.90

Heat Flow (Btu/hr) -1.694e+008 -1.694e+008

Liq Vol Flow @Std Cond (barrel/day) 4810 4810

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	31.77	31.77
Molar Density (lbmole/ft3)	1.304	1.304
Mass Density (lb/ft3)	41.43	41.43
Act. Volume Flow (barrel/day)	5467	5467
Mass Enthalpy (Btu/lb)	-3196	-3196
Mass Entropy (Btu/lb-F)	0.9095	0.9095
Heat Capacity (Btu/lbmole-F)	28.21	28.21
Mass Heat Capacity (Btu/lb-F)	0.8880	0.8880
LHV Molar Basis (Std) (Btu/lbmole)	2.768e+005	2.768e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.117e+005	3.117e+005
HHV Mass Basis (Std) (Btu/lb)	9812	9812
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8713	8713
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-113.5	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	1.486	1.486
Specific Heat (Btu/lbmole-F)	28.21	28.21
Std. Gas Flow (MMSCFD)	15.16	15.16
Std. Ideal Liq. Mass Density (lb/ft3)	47.22	47.22
Act. Liq. Flow (USGPS)	2.658	2.658
Z Factor	5.400e-002	5.400e-002
Watson K	11.19	11.19
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.076	1.076
Cp/Cv	1.248	1.248
Heat of Vap. (Btu/lbmole)	1.256e+004	---
Kinematic Viscosity (cSt)	0.2062	0.2062
Liq. Mass Density (Std. Cond) (lb/ft3)	47.10	47.10
Liq. Vol. Flow (Std. Cond) (barrel/day)	4810	4810
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.7668	0.7668
Mass Heat of Vap. (Btu/lb)	395.5	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	28.24	28.24
Thermal Conductivity (Btu/hr-ft-F)	0.1408	0.1408

Viscosity (cP)	0.1368	0.1368
Cv (Semi-Ideal) (Btu/lbmole-F)	26.23	26.23
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.8255	0.8255
Cv (Btu/lbmole-F)	22.60	22.60
Mass Cv (Btu/lb-F)	0.7114	0.7114
Cv (Ent. Method) (Btu/lbmole-F)	124.2	124.2
Mass Cv (Ent. Method) (Btu/lb-F)	3.908	3.908
Cp/Cv (Ent. Method)	0.2272	0.2272
Reid VP at 37.8 C (psia)	65.93	65.93
True VP at 37.8 C (psia)	40.34	40.34
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	4810	4810
Viscosity Index	-63.00	---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	610.1179	0.3658	10991.3349	0.2074	754.1266	0.1572
diM-Ether	577.8707	0.3464	26621.9235	0.5024	2719.4883	0.5669
Methanol	480.0121	0.2878	15380.4997	0.2902	1323.5138	0.2759
Total	1668.0006	1.0000	52993.7581	1.0000	4797.1286	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	610.1179	0.3658	10991.3349	0.2074	754.1266	0.1572
diM-Ether	577.8707	0.3464	26621.9235	0.5024	2719.4883	0.5669
Methanol	480.0121	0.2878	15380.4997	0.2902	1323.5138	0.2759
Total	1668.0006	1.0000	52993.7581	1.0000	4797.1286	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Distillation: T-100	Cooler: E-101	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: DME Outlet Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	117.3	117.3
Pressure: (psia)	153.7	153.7
Molar Flow (lbmole/hr)	578.0	578.0
Mass Flow (lb/hr)	2.662e+004	2.662e+004
Std Ideal Liq Vol Flow (barrel/day)	2719	2719
Molar Enthalpy (Btu/lbmole)	-8.641e+004	-8.641e+004
Molar Entropy (Btu/lbmole-F)	38.59	38.59
Heat Flow (Btu/hr)	-4.995e+007	-4.995e+007
Liq Vol Flow @Std Cond (barrel/day)	2950	2950

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	46.06	46.06
Molar Density (lbmole/ft3)	0.7709	0.7709
Mass Density (lb/ft3)	35.51	35.51
Act. Volume Flow (barrel/day)	3205	3205
Mass Enthalpy (Btu/lb)	-1876	-1876
Mass Entropy (Btu/lb-F)	0.8378	0.8378
Heat Capacity (Btu/lbmole-F)	28.61	28.61
Mass Heat Capacity (Btu/lb-F)	0.6212	0.6212
LHV Molar Basis (Std) (Btu/lbmole)	5.709e+005	5.709e+005
HHV Molar Basis (Std) (Btu/lbmole)	6.238e+005	6.238e+005
HHV Mass Basis (Std) (Btu/lb)	1.354e+004	1.354e+004
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	1.240e+004	1.240e+004
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-65.72	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	0.9086	0.9086
Specific Heat (Btu/lbmole-F)	28.61	28.61
Std. Gas Flow (MMSCFD)	5.254	5.254
Std. Ideal Liq. Mass Density (lb/ft3)	41.85	41.85
Act. Liq. Flow (USGPS)	1.558	1.558
Z Factor	3.221e-002	3.221e-002
Watson K	11.38	11.38

User Property --- ---

Partial Pressure of H2S (psia) 0.0000 ---

Cp/(Cp - R) 1.075 1.075

Cp/Cv 1.075 1.075

Heat of Vap. (Btu/lbmole) 7879 ---

Kinematic Viscosity (cSt) 0.1278 0.1278

Liq. Mass Density (Std. Cond) (lb/ft3) 38.58 38.58

Liq. Vol. Flow (Std. Cond) (barrel/day) 2950 2950

Liquid Fraction 1.000 1.000

Molar Volume (ft3/lbmole) 1.297 1.297

Mass Heat of Vap. (Btu/lb) 171.1 ---

Phase Fraction [Molar Basis] 0.0000 1.0000

Surface Tension (dyne/cm) 8.712 8.712

Thermal Conductivity (Btu/hr-ft-F) 6.851e-002 6.851e-002

Viscosity (cP) 7.266e-002 7.266e-002

Cv (Semi-Ideal) (Btu/lbmole-F) 26.62 26.62

Mass Cv (Semi-Ideal) (Btu/lb-F) 0.5780 0.5780

Cv (Btu/lbmole-F) 26.62 26.62

Mass Cv (Btu/lb-F) 0.5780 0.5780

Cv (Ent. Method) (Btu/lbmole-F) --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- ---

Cp/Cv (Ent. Method) --- ---

Reid VP at 37.8 C (psia) 120.5 120.5

True VP at 37.8 C (psia) 120.5 120.5

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 2950 2950

Viscosity Index --- ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0055	0.0000	0.0996	0.0000	0.0068	0.0000
diM-Ether	577.6286	0.9993	26610.7715	0.9995	2718.3491	0.9996
Methanol	0.4123	0.0007	13.2116	0.0005	1.1369	0.0004
Total	578.0464	1.0000	26624.0827	1.0000	2719.4928	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0055	0.0000	0.0996	0.0000	0.0068	0.0000
diM-Ether	577.6286	0.9993	26610.7715	0.9995	2718.3491	0.9996
Methanol	0.4123	0.0007	13.2116	0.0005	1.1369	0.0004
Total	578.0464	1.0000	26624.0827	1.0000	2719.4928	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---

diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Distillation: T-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Methanol and Water Outlet Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Aqueous Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	305.2	305.2
Pressure: (psia)	156.0	156.0
Molar Flow (lbmole/hr)	1090	1090
Mass Flow (lb/hr)	2.637e+004	2.637e+004
Std Ideal Liq Vol Flow (barrel/day)	2078	2078
Molar Enthalpy (Btu/lbmole)	-1.085e+005	-1.085e+005
Molar Entropy (Btu/lbmole-F)	21.07	21.07
Heat Flow (Btu/hr)	-1.183e+008	-1.183e+008
Liq Vol Flow @Std Cond (barrel/day)	2058	2058

PROPERTIES

	Overall	Aqueous Phase
Molecular Weight	24.19	24.19
Molar Density (lbmole/ft3)	1.907	1.907
Mass Density (lb/ft3)	46.14	46.14
Act. Volume Flow (barrel/day)	2443	2443
Mass Enthalpy (Btu/lb)	-4486	-4486
Mass Entropy (Btu/lb-F)	0.8709	0.8709
Heat Capacity (Btu/lbmole-F)	25.61	25.61
Mass Heat Capacity (Btu/lb-F)	1.059	1.059
LHV Molar Basis (Std) (Btu/lbmole)	1.208e+005	1.208e+005
HHV Molar Basis (Std) (Btu/lbmole)	1.462e+005	1.462e+005
HHV Mass Basis (Std) (Btu/lb)	6044	6044
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	4995	4995

Phase Fraction [Vol. Basis] 0.0000 1.000

Phase Fraction [Mass Basis] 0.0000 1.000

Phase Fraction [Act. Vol. Basis] 0.0000 1.000

Mass Exergy (Btu/lb) -87.38 ---

Partial Pressure of CO2 (psia) 0.0000 ---

Cost Based on Flow (Cost/s) 0.0000 0.0000

Act. Gas Flow (ACFM) --- ---

Avg. Liq. Density (lbmole/ft3) 2.243 2.243

Specific Heat (Btu/lbmole-F) 25.61 25.61

Std. Gas Flow (MMSCFD) 9.908 9.908

Std. Ideal Liq. Mass Density (lb/ft3) 54.25 54.25

Act. Liq. Flow (USGPS) 1.188 1.188

Z Factor 9.966e-003 9.966e-003

Watson K 10.63 10.63

User Property --- ---

Partial Pressure of H2S (psia) 0.0000 ---

Cp/(Cp - R) 1.084 1.084

Cp/Cv 1.233 1.233

Heat of Vap. (Btu/lbmole) 1.474e+004 ---

Kinematic Viscosity (cSt) 0.1957 0.1957

Liq. Mass Density (Std. Cond) (lb/ft3) 54.76 54.76

Liq. Vol. Flow (Std. Cond) (barrel/day) 2058 2058

Liquid Fraction 1.000 1.000

Molar Volume (ft3/lbmole) 0.5244 0.5244

Mass Heat of Vap. (Btu/lb) 609.1 ---

Phase Fraction [Molar Basis] 0.0000 1.0000

Surface Tension (dyne/cm) 31.66 31.66

Thermal Conductivity (Btu/hr-ft-F) 0.2087 0.2087

Viscosity (cP) 0.1447 0.1447

Cv (Semi-Ideal) (Btu/lbmole-F) 23.62 23.62

Mass Cv (Semi-Ideal) (Btu/lb-F) 0.9765 0.9765

Cv (Btu/lbmole-F) 20.78 20.78

Mass Cv (Btu/lb-F) 0.8588 0.8588

Cv (Ent. Method) (Btu/lbmole-F) --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- ---

Cp/Cv (Ent. Method) --- ---

Reid VP at 37.8 C (psia) 4.697 4.697

True VP at 37.8 C (psia) 3.173 3.173

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 2058 2058

Viscosity Index --- ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	610.1124	0.5598	10991.2353	0.4168	754.1198	0.3630
diM-Ether	0.2421	0.0002	11.1519	0.0004	1.1392	0.0005

Methanol	479.5998	0.4400	15367.2881	0.5828	1322.3769	0.6365
Total	1089.9542	1.0000	26369.6753	1.0000	2077.6358	1.0000

Aqueous Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	610.1124	0.5598	10991.2353	0.4168	754.1198	0.3630
diM-Ether	0.2421	0.0002	11.1519	0.0004	1.1392	0.0005
Methanol	479.5998	0.4400	15367.2881	0.5828	1322.3769	0.6365
Total	1089.9542	1.0000	26369.6753	1.0000	2077.6358	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	---	0.0000
diM-Ether	0.0000	---	0.0000
Methanol	0.0000	---	0.0000

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Distillation: T-101	Distillation: T-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Wastewater Out Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Vapour Phase	Aqueous Phase
Vapour / Phase Fraction	0.0000	0.0000	1.0000
Temperature: (F)	344.8	344.8	344.8
Pressure: (psia)	128.4	128.4	128.4
Molar Flow (lbmole/hr)	582.2	1.551e-003	582.2
Mass Flow (lb/hr)	1.053e+004	2.848e-002	1.053e+004
Std Ideal Liq Vol Flow (barrel/day)	724.1	1.976e-003	724.1
Molar Enthalpy (Btu/lbmole)	-1.175e+005	-1.014e+005	-1.175e+005
Molar Entropy (Btu/lbmole-F)	9.146	29.14	9.146
Heat Flow (Btu/hr)	-6.840e+007	-157.2	-6.840e+007
Liq Vol Flow @Std Cond (barrel/day)	712.1	1.943e-003	712.1

PROPERTIES

	Overall	Vapour Phase	Aqueous Phase
Molecular Weight	18.09	18.37	18.09

Molar Density (lbmole/ft3)	3.030	1.487e-002	3.031
Mass Density (lb/ft3)	54.80	0.2732	54.82
Act. Volume Flow (barrel/day)	821.4	0.4457	821.0
Mass Enthalpy (Btu/lb)	-6496	-5518	-6496
Mass Entropy (Btu/lb-F)	0.5057	1.586	0.5057
Heat Capacity (Btu/lbmole-F)	19.01	12.28	19.01
Mass Heat Capacity (Btu/lb-F)	1.051	0.6684	1.051
LHV Molar Basis (Std) (Btu/lbmole)	1372	6893	1372
HHV Molar Basis (Std) (Btu/lbmole)	1.909e+004	2.497e+004	1.909e+004
HHV Mass Basis (Std) (Btu/lb)	1056	1359	1056
CO2 Loading	---	---	---
CO2 App ML Con (lbmole/ft3)	---	---	---
CO2 App WT Con (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	75.85	375.3	75.85
Phase Fraction [Vol. Basis]	2.729e-006	2.729e-006	1.000
Phase Fraction [Mass Basis]	2.705e-006	2.705e-006	1.000
Phase Fraction [Act. Vol. Basis]	5.426e-004	5.426e-004	0.9995
Mass Exergy (Btu/lb)	49.92	---	---
Partial Pressure of CO2 (psia)	0.0000	---	---
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	1.738e-003	1.738e-003	---
Avg. Liq. Density (lbmole/ft3)	3.437	3.355	3.437
Specific Heat (Btu/lbmole-F)	19.01	12.28	19.01
Std. Gas Flow (MMSCFD)	5.293	1.410e-005	5.293
Std. Ideal Liq. Mass Density (lb/ft3)	62.16	61.62	62.16
Act. Liq. Flow (USGPS)	0.3991	---	0.3991
Z Factor	---	1.000	4.906e-003
Watson K	10.63	10.63	10.63
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	---	---
Cp/(Cp - R)	1.117	1.193	1.117
Cp/Cv	1.115	1.193	1.213
Heat of Vap. (Btu/lbmole)	1.575e+004	---	---
Kinematic Viscosity (cSt)	---	2.591	0.1026
Liq. Mass Density (Std. Cond) (lb/ft3)	63.21	62.66	63.21
Liq. Vol. Flow (Std. Cond) (barrel/day)	712.1	1.943e-003	712.1
Liquid Fraction	1.000	0.0000	1.000
Molar Volume (ft3/lbmole)	0.3301	67.24	0.3299
Mass Heat of Vap. (Btu/lb)	870.9	---	---
Phase Fraction [Molar Basis]	0.0000	0.0000	1.0000
Surface Tension (dyne/cm)	43.44	---	43.44
Thermal Conductivity (Btu/hr-ft-F)	0.3918	1.826e-002	0.3918
Viscosity (cP)	9.011e-002	1.134e-002	9.011e-002
Cv (Semi-Ideal) (Btu/lbmole-F)	17.02	10.29	17.02
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.9413	0.5602	0.9413
Cv (Btu/lbmole-F)	17.04	10.29	15.67
Mass Cv (Btu/lb-F)	0.9424	0.5602	0.8662
Cv (Ent. Method) (Btu/lbmole-F)	---	---	---

Mass Cv (Ent. Method) (Btu/lb-F) --- --- ---

Cp/Cv (Ent. Method) --- --- ---

Reid VP at 37.8 C (psia) 4.641 4.641 4.641

True VP at 37.8 C (psia) 1.003 1.199 1.003

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 712.1 1.943e-003 712.1

Viscosity Index --- --- ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	579.3235	0.9950	10436.5702	0.9911	716.0636	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	2.9112	0.0050	93.2813	0.0089	8.0270	0.0111
Total	582.2347	1.0000	10529.8515	1.0000	724.0906	1.0000

Vapour Phase Phase Fraction 2.663e-006

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0015	0.9749	0.0272	0.9562	0.0019	0.9456
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0251	0.0012	0.0438	0.0001	0.0544
Total	0.0016	1.0000	0.0285	1.0000	0.0020	1.0000

Aqueous Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	579.3219	0.9950	10436.5430	0.9911	716.0618	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	2.9112	0.0050	93.2800	0.0089	8.0269	0.0111
Total	582.2331	1.0000	10529.8230	1.0000	724.0886	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.9798	---	0.9798
diM-Ether	11.73	---	11.73
Methanol	5.025	---	5.025

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Separator: V-100	Distillation: T-101	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Methanol Out | Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	269.2	269.2
Pressure: (psia)	124.7	124.7
Molar Flow (lbmole/hr)	507.7	507.7
Mass Flow (lb/hr)	1.584e+004	1.584e+004
Std Ideal Liq Vol Flow (barrel/day)	1354	1354
Molar Enthalpy (Btu/lbmole)	-9.858e+004	-9.858e+004
Molar Entropy (Btu/lbmole-F)	30.69	30.69
Heat Flow (Btu/hr)	-5.005e+007	-5.005e+007
Liq Vol Flow @Std Cond (barrel/day)	1352	1352

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	31.20	31.20
Molar Density (lbmole/ft3)	1.334	1.334
Mass Density (lb/ft3)	41.63	41.63
Act. Volume Flow (barrel/day)	1626	1626
Mass Enthalpy (Btu/lb)	-3160	-3160
Mass Entropy (Btu/lb-F)	0.9836	0.9836
Heat Capacity (Btu/lbmole-F)	31.67	31.67
Mass Heat Capacity (Btu/lb-F)	1.015	1.015
LHV Molar Basis (Std) (Btu/lbmole)	2.578e+005	2.578e+005
HHV Molar Basis (Std) (Btu/lbmole)	2.920e+005	2.920e+005
HHV Mass Basis (Std) (Btu/lb)	9361	9361
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8265	8265
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-160.4	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	1.603	1.603
Specific Heat (Btu/lbmole-F)	31.67	31.67
Std. Gas Flow (MMSCFD)	4.615	4.615
Std. Ideal Liq. Mass Density (lb/ft3)	50.02	50.02
Act. Liq. Flow (USGPS)	0.7906	0.7906

Z Factor	1.195e-002	1.195e-002
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.067	1.067
Cp/Cv	1.381	1.381
Heat of Vap. (Btu/lbmole)	1.320e+004	---
Kinematic Viscosity (cSt)	0.2437	0.2437
Liq. Mass Density (Std. Cond) (lb/ft3)	50.08	50.08
Liq. Vol. Flow (Std. Cond) (barrel/day)	1352	1352
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.7494	0.7494
Mass Heat of Vap. (Btu/lb)	423.0	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	15.68	15.68
Thermal Conductivity (Btu/hr-ft-F)	8.659e-002	8.659e-002
Viscosity (cP)	0.1625	0.1625
Cv (Semi-Ideal) (Btu/lbmole-F)	29.68	29.68
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.9513	0.9513
Cv (Btu/lbmole-F)	22.93	22.93
Mass Cv (Btu/lb-F)	0.7349	0.7349
Cv (Ent. Method) (Btu/lbmole-F)	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---
Cp/Cv (Ent. Method)	---	---
Reid VP at 37.8 C (psia)	4.697	4.697
True VP at 37.8 C (psia)	4.525	4.525
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	1352	1352
Viscosity Index	-34.84	---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	30.7889	0.0606	554.6651	0.0350	38.0561	0.0281
diM-Ether	0.2421	0.0005	11.1519	0.0007	1.1392	0.0008
Methanol	476.6885	0.9389	15274.0068	0.9643	1314.3499	0.9710
Total	507.7195	1.0000	15839.8238	1.0000	1353.5452	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	30.7889	0.0606	554.6651	0.0350	38.0561	0.0281
diM-Ether	0.2421	0.0005	11.1519	0.0007	1.1392	0.0008
Methanol	476.6885	0.9389	15274.0068	0.9643	1314.3499	0.9710
Total	507.7195	1.0000	15839.8238	1.0000	1353.5452	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Mixer: MIX-101	Distillation: T-101	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Methanol Mix Feed Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	137.6	137.6
Pressure: (psia)	14.65	14.65
Molar Flow (lbmole/hr)	1668	1668
Mass Flow (lb/hr)	5.299e+004	5.299e+004
Std Ideal Liq Vol Flow (barrel/day)	4550	4550
Molar Enthalpy (Btu/lbmole)	-1.016e+005	-1.016e+005
Molar Entropy (Btu/lbmole-F)	17.08	17.08
Heat Flow (Btu/hr)	-1.695e+008	-1.695e+008
Liq Vol Flow @Std Cond (barrel/day)	4546	4546

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	31.77	31.77
Molar Density (lbmole/ft3)	1.477	1.477
Mass Density (lb/ft3)	46.93	46.93
Act. Volume Flow (barrel/day)	4827	4827
Mass Enthalpy (Btu/lb)	-3199	-3199
Mass Entropy (Btu/lb-F)	0.5375	0.5375
Heat Capacity (Btu/lbmole-F)	28.31	28.31
Mass Heat Capacity (Btu/lb-F)	0.8911	0.8911
LHV Molar Basis (Std) (Btu/lbmole)	2.690e+005	2.690e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.040e+005	3.040e+005
HHV Mass Basis (Std) (Btu/lb)	9567	9567
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---

CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8468	8468
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-46.84	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	1.567	1.567
Specific Heat (Btu/lbmole-F)	28.31	28.31
Std. Gas Flow (MMSCFD)	15.16	15.16
Std. Ideal Liq. Mass Density (lb/ft3)	49.78	49.78
Act. Liq. Flow (USGPS)	2.346	2.346
Z Factor	1.547e-003	1.547e-003
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.075	1.075
Cp/Cv	1.373	1.373
Heat of Vap. (Btu/lbmole)	1.524e+004	---
Kinematic Viscosity (cSt)	0.4719	0.4719
Liq. Mass Density (Std. Cond) (lb/ft3)	49.83	49.83
Liq. Vol. Flow (Std. Cond) (barrel/day)	4546	4546
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.6770	0.6770
Mass Heat of Vap. (Btu/lb)	479.5	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	25.10	25.10
Thermal Conductivity (Btu/hr-ft-F)	9.889e-002	9.889e-002
Viscosity (cP)	0.3548	0.3548
Cv (Semi-Ideal) (Btu/lbmole-F)	26.33	26.33
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.8286	0.8286
Cv (Btu/lbmole-F)	20.61	20.61
Mass Cv (Btu/lb-F)	0.6488	0.6488
Cv (Ent. Method) (Btu/lbmole-F)	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---
Cp/Cv (Ent. Method)	---	---
Reid VP at 37.8 C (psia)	4.669	4.669
True VP at 37.8 C (psia)	4.615	4.615
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	4546	4546
Viscosity Index	-9.738	---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS MOLAR FLOW MOLE FRACTION MASS FLOW MASS FRACTION LIQUID VOLUME LIQUID VOLUME

(lbmole/hr) (lb/hr) FLOW (barrel/day) FRACTION

H2O	32.6646	0.0196	588.4568	0.0111	40.3746	0.0089
diM-Ether	0.4174	0.0003	19.2302	0.0004	1.9644	0.0004
Methanol	1634.9186	0.9802	52385.8982	0.9885	4507.8807	0.9907
Total	1668.0006	1.0000	52993.5852	1.0000	4550.2197	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	32.6646	0.0196	588.4568	0.0111	40.3746	0.0089
diM-Ether	0.4174	0.0003	19.2302	0.0004	1.9644	0.0004
Methanol	1634.9186	0.9802	52385.8982	0.9885	4507.8807	0.9907
Total	1668.0006	1.0000	52993.5852	1.0000	4550.2197	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Pump: P-100	Mixer: MIX-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Methanol Recycle Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	269.0	269.0
Pressure: (psia)	124.7	124.7
Molar Flow (lbmole/hr)	507.7	507.7
Mass Flow (lb/hr)	1.584e+004	1.584e+004
Std Ideal Liq Vol Flow (barrel/day)	1354	1354
Molar Enthalpy (Btu/lbmole)	-9.857e+004	-9.857e+004
Molar Entropy (Btu/lbmole-F)	30.70	30.70
Heat Flow (Btu/hr)	-5.004e+007	-5.004e+007
Liq Vol Flow @Std Cond (barrel/day)	1353	1353

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	31.21	31.21
Molar Density (lbmole/ft3)	1.334	1.334
Mass Density (lb/ft3)	41.63	41.63
Act. Volume Flow (barrel/day)	1627	1627
Mass Enthalpy (Btu/lb)	-3158	-3158
Mass Entropy (Btu/lb-F)	0.9836	0.9836
Heat Capacity (Btu/lbmole-F)	31.66	31.66
Mass Heat Capacity (Btu/lb-F)	1.015	1.015
LHV Molar Basis (Std) (Btu/lbmole)	2.580e+005	2.580e+005
HHV Molar Basis (Std) (Btu/lbmole)	2.923e+005	2.923e+005
HHV Mass Basis (Std) (Btu/lb)	9365	9365
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8268	8268
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-160.3	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	1.603	1.603
Specific Heat (Btu/lbmole-F)	31.66	31.66
Std. Gas Flow (MMSCFD)	4.615	4.615
Std. Ideal Liq. Mass Density (lb/ft3)	50.02	50.02
Act. Liq. Flow (USGPS)	0.7909	0.7909
Z Factor	1.196e-002	1.196e-002
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.067	1.067
Cp/Cv	1.382	1.382
Heat of Vap. (Btu/lbmole)	1.320e+004	---
Kinematic Viscosity (cSt)	0.2437	0.2437
Liq. Mass Density (Std. Cond) (lb/ft3)	50.07	50.07
Liq. Vol. Flow (Std. Cond) (barrel/day)	1353	1353
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.7497	0.7497
Mass Heat of Vap. (Btu/lb)	422.8	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	15.67	15.67
Thermal Conductivity (Btu/hr-ft-F)	8.651e-002	8.651e-002
Viscosity (cP)	0.1625	0.1625
Cv (Semi-Ideal) (Btu/lbmole-F)	29.68	29.68
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.9509	0.9509
Cv (Btu/lbmole-F)	22.92	22.92

Mass Cv (Btu/lb-F) 0.7343 0.7343

Cv (Ent. Method) (Btu/lbmole-F) --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- ---

Cp/Cv (Ent. Method) --- ---

Reid VP at 37.8 C (psia) 4.737 4.737

True VP at 37.8 C (psia) 4.565 4.565

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 1353 1353

Viscosity Index -34.83 ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	30.6026	0.0603	551.3081	0.0348	37.8258	0.0279
diM-Ether	0.4174	0.0008	19.2302	0.0012	1.9644	0.0015
Methanol	476.6997	0.9389	15274.3635	0.9640	1314.3806	0.9706
Total	507.7196	1.0000	15844.9018	1.0000	1354.1708	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	30.6026	0.0603	551.3081	0.0348	37.8258	0.0279
diM-Ether	0.4174	0.0008	19.2302	0.0012	1.9644	0.0015
Methanol	476.6997	0.9389	15274.3635	0.9640	1314.3806	0.9706
Total	507.7196	1.0000	15844.9018	1.0000	1354.1708	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Mixer: MIX-100	Recycle: RCY-1	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Pressurized Methanol Feed Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	140.4	140.4
Pressure: (psia)	500.0	500.0
Molar Flow (lbmole/hr)	1668	1668
Mass Flow (lb/hr)	5.299e+004	5.299e+004
Std Ideal Liq Vol Flow (barrel/day)	4550	4550
Molar Enthalpy (Btu/lbmole)	-1.016e+005	-1.016e+005
Molar Entropy (Btu/lbmole-F)	24.20	24.20
Heat Flow (Btu/hr)	-1.694e+008	-1.694e+008
Liq Vol Flow @Std Cond (barrel/day)	4546	4546

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	31.77	31.77
Molar Density (lbmole/ft3)	1.478	1.478
Mass Density (lb/ft3)	46.97	46.97
Act. Volume Flow (barrel/day)	4823	4823
Mass Enthalpy (Btu/lb)	-3196	-3196
Mass Entropy (Btu/lb-F)	0.7619	0.7619
Heat Capacity (Btu/lbmole-F)	28.36	28.36
Mass Heat Capacity (Btu/lb-F)	0.8927	0.8927
LHV Molar Basis (Std) (Btu/lbmole)	2.690e+005	2.690e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.040e+005	3.040e+005
HHV Mass Basis (Std) (Btu/lb)	9567	9567
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8468	8468
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-164.7	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	1.567	1.567
Specific Heat (Btu/lbmole-F)	28.36	28.36
Std. Gas Flow (MMSCFD)	15.16	15.16
Std. Ideal Liq. Mass Density (lb/ft3)	49.78	49.78
Act. Liq. Flow (USGPS)	2.344	2.344
Z Factor	5.252e-002	5.252e-002
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.075	1.075
Cp/Cv	1.365	1.365

Heat of Vap. (Btu/lbmole)	1.050e+004	---
Kinematic Viscosity (cSt)	0.4637	0.4637
Liq. Mass Density (Std. Cond) (lb/ft3)	49.83	49.83
Liq. Vol. Flow (Std. Cond) (barrel/day)	4546	4546
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.6764	0.6764
Mass Heat of Vap. (Btu/lb)	330.6	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	24.85	24.85
Thermal Conductivity (Btu/hr-ft-F)	9.850e-002	9.850e-002
Viscosity (cP)	0.3489	0.3489
Cv (Semi-Ideal) (Btu/lbmole-F)	26.38	26.38
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.8302	0.8302
Cv (Btu/lbmole-F)	20.79	20.79
Mass Cv (Btu/lb-F)	0.6542	0.6542
Cv (Ent. Method) (Btu/lbmole-F)	337.6	337.6
Mass Cv (Ent. Method) (Btu/lb-F)	10.63	10.63
Cp/Cv (Ent. Method)	8.402e-002	8.402e-002
Reid VP at 37.8 C (psia)	4.669	4.669
True VP at 37.8 C (psia)	4.615	4.615
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	4546	4546
Viscosity Index	-10.14	---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	32.6646	0.0196	588.4568	0.0111	40.3746	0.0089
diM-Ether	0.4174	0.0003	19.2302	0.0004	1.9644	0.0004
Methanol	1634.9186	0.9802	52385.8982	0.9885	4507.8807	0.9907
Total	1668.0006	1.0000	52993.5852	1.0000	4550.2197	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	32.6646	0.0196	588.4568	0.0111	40.3746	0.0089
diM-Ether	0.4174	0.0003	19.2302	0.0004	1.9644	0.0004
Methanol	1634.9186	0.9802	52385.8982	0.9885	4507.8807	0.9907
Total	1668.0006	1.0000	52993.5852	1.0000	4550.2197	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Heater: E-100	Pump: P-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Mixed MeOH   Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	269.2	269.2
Pressure: (psia)	124.7	124.7
Molar Flow (lbmole/hr)	507.7	507.7
Mass Flow (lb/hr)	1.584e+004	1.584e+004
Std Ideal Liq Vol Flow (barrel/day)	1354	1354
Molar Enthalpy (Btu/lbmole)	-9.858e+004	-9.858e+004
Molar Entropy (Btu/lbmole-F)	30.69	30.69
Heat Flow (Btu/hr)	-5.005e+007	-5.005e+007
Liq Vol Flow @Std Cond (barrel/day)	1352	1352

PROPERTIES

	Overall	Liquid Phase
Molecular Weight	31.20	31.20
Molar Density (lbmole/ft3)	1.334	1.334
Mass Density (lb/ft3)	41.63	41.63
Act. Volume Flow (barrel/day)	1626	1626
Mass Enthalpy (Btu/lb)	-3160	-3160
Mass Entropy (Btu/lb-F)	0.9836	0.9836
Heat Capacity (Btu/lbmole-F)	31.67	31.67
Mass Heat Capacity (Btu/lb-F)	1.015	1.015
LHV Molar Basis (Std) (Btu/lbmole)	2.578e+005	2.578e+005
HHV Molar Basis (Std) (Btu/lbmole)	2.920e+005	2.920e+005
HHV Mass Basis (Std) (Btu/lb)	9361	9361
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	8265	8265
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	-160.4	---

Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	1.603	1.603
Specific Heat (Btu/lbmole-F)	31.67	31.67
Std. Gas Flow (MMSCFD)	4.615	4.615
Std. Ideal Liq. Mass Density (lb/ft3)	50.02	50.02
Act. Liq. Flow (USGPS)	0.7906	0.7906
Z Factor	1.195e-002	1.195e-002
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.067	1.067
Cp/Cv	1.381	1.381
Heat of Vap. (Btu/lbmole)	1.320e+004	---
Kinematic Viscosity (cSt)	0.2437	0.2437
Liq. Mass Density (Std. Cond) (lb/ft3)	50.08	50.08
Liq. Vol. Flow (Std. Cond) (barrel/day)	1352	1352
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.7494	0.7494
Mass Heat of Vap. (Btu/lb)	423.0	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	15.68	15.68
Thermal Conductivity (Btu/hr-ft-F)	8.659e-002	8.659e-002
Viscosity (cP)	0.1625	0.1625
Cv (Semi-Ideal) (Btu/lbmole-F)	29.68	29.68
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.9513	0.9513
Cv (Btu/lbmole-F)	22.93	22.93
Mass Cv (Btu/lb-F)	0.7349	0.7349
Cv (Ent. Method) (Btu/lbmole-F)	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---
Cp/Cv (Ent. Method)	---	---
Reid VP at 37.8 C (psia)	4.697	4.697
True VP at 37.8 C (psia)	4.525	4.525
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	1352	1352
Viscosity Index	-34.84	---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	30.7889	0.0606	554.6651	0.0350	38.0561	0.0281
diM-Ether	0.2421	0.0005	11.1519	0.0007	1.1392	0.0008
Methanol	476.6885	0.9389	15274.0068	0.9643	1314.3499	0.9710
Total	507.7195	1.0000	15839.8238	1.0000	1353.5452	1.0000

Liquid Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	30.7889	0.0606	554.6651	0.0350	38.0561	0.0281
diM-Ether	0.2421	0.0005	11.1519	0.0007	1.1392	0.0008
Methanol	476.6885	0.9389	15274.0068	0.9643	1314.3499	0.9710
Total	507.7195	1.0000	15839.8238	1.0000	1353.5452	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	0.0000	---
diM-Ether	0.0000	0.0000	---
Methanol	0.0000	0.0000	---

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Recycle: RCY-1	Mixer: MIX-101	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Vapor Check   Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Vapour Phase	Aqueous Phase
Vapour / Phase Fraction	1.0000	1.0000	0.0000
Temperature: (F)	344.8	344.8	344.8
Pressure: (psia)	128.4	128.4	128.4
Molar Flow (lbmole/hr)	1.551e-003	1.551e-003	0.0000
Mass Flow (lb/hr)	2.848e-002	2.848e-002	0.0000
Std Ideal Liq Vol Flow (barrel/day)	1.976e-003	1.976e-003	0.0000
Molar Enthalpy (Btu/lbmole)	-1.014e+005	-1.014e+005	-1.175e+005
Molar Entropy (Btu/lbmole-F)	29.14	29.14	9.146
Heat Flow (Btu/hr)	-157.2	-157.2	0.0000
Liq Vol Flow @Std Cond (barrel/day)	1.943e-003	1.943e-003	0.0000

PROPERTIES

	Overall	Vapour Phase	Aqueous Phase
Molecular Weight	18.37	18.37	18.09
Molar Density (lbmole/ft3)	1.487e-002	1.487e-002	3.031
Mass Density (lb/ft3)	0.2732	0.2732	54.82
Act. Volume Flow (barrel/day)	0.4457	0.4457	0.0000
Mass Enthalpy (Btu/lb)	-5518	-5518	-6496

Mass Entropy (Btu/lb-F)	1.586	1.586	0.5057
Heat Capacity (Btu/lbmole-F)	12.28	12.28	19.01
Mass Heat Capacity (Btu/lb-F)	0.6684	0.6684	1.051
LHV Molar Basis (Std) (Btu/lbmole)	6893	6893	1372
HHV Molar Basis (Std) (Btu/lbmole)	2.497e+004	2.497e+004	1.909e+004
HHV Mass Basis (Std) (Btu/lb)	1359	1359	1056
CO2 Loading	---	---	---
CO2 App ML Con (lbmole/ft3)	---	---	---
CO2 App WT Con (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	375.3	375.3	75.85
Phase Fraction [Vol. Basis]	1.000	1.000	---
Phase Fraction [Mass Basis]	1.000	1.000	0.0000
Phase Fraction [Act. Vol. Basis]	1.000	1.000	0.0000
Mass Exergy (Btu/lb)	331.8	---	---
Partial Pressure of CO2 (psia)	0.0000	---	---
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	1.738e-003	1.738e-003	---
Avg. Liq. Density (lbmole/ft3)	3.355	3.355	3.437
Specific Heat (Btu/lbmole-F)	12.28	12.28	19.01
Std. Gas Flow (MMSCFD)	1.410e-005	1.410e-005	0.0000
Std. Ideal Liq. Mass Density (lb/ft3)	61.62	61.62	62.16
Act. Liq. Flow (USGPS)	0.0000	---	0.0000
Z Factor	---	1.000	4.906e-003
Watson K	10.63	10.63	10.63
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	---	---
Cp/(Cp - R)	1.193	1.193	1.117
Cp/Cv	1.193	1.193	1.213
Heat of Vap. (Btu/lbmole)	1.577e+004	---	---
Kinematic Viscosity (cSt)	2.591	2.591	0.1026
Liq. Mass Density (Std. Cond) (lb/ft3)	62.66	62.66	63.21
Liq. Vol. Flow (Std. Cond) (barrel/day)	1.943e-003	1.943e-003	0.0000
Liquid Fraction	0.0000	0.0000	1.000
Molar Volume (ft3/lbmole)	67.24	67.24	0.3299
Mass Heat of Vap. (Btu/lb)	858.4	---	---
Phase Fraction [Molar Basis]	1.0000	1.0000	0.0000
Surface Tension (dyne/cm)	---	---	43.44
Thermal Conductivity (Btu/hr-ft-F)	1.826e-002	1.826e-002	0.3918
Viscosity (cP)	1.134e-002	1.134e-002	9.011e-002
Cv (Semi-Ideal) (Btu/lbmole-F)	10.29	10.29	17.02
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.5602	0.5602	0.9413
Cv (Btu/lbmole-F)	10.29	10.29	15.67
Mass Cv (Btu/lb-F)	0.5602	0.5602	0.8662
Cv (Ent. Method) (Btu/lbmole-F)	---	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---	---
Cp/Cv (Ent. Method)	---	---	---
Reid VP at 37.8 C (psia)	4.641	4.641	4.641
True VP at 37.8 C (psia)	1.199	1.199	1.003

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 1.943e-003 1.943e-003 0.0000

Viscosity Index --- --- ---

COMPOSITION

Overall Phase Vapour Fraction 1.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0015	0.9749	0.0272	0.9562	0.0019	0.9456
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0251	0.0012	0.0438	0.0001	0.0544
Total	0.0016	1.0000	0.0285	1.0000	0.0020	1.0000

Vapour Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0015	0.9749	0.0272	0.9562	0.0019	0.9456
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0251	0.0012	0.0438	0.0001	0.0544
Total	0.0016	1.0000	0.0285	1.0000	0.0020	1.0000

Aqueous Phase Phase Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0000	0.9950	0.0000	0.9911	0.0000	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0050	0.0000	0.0089	0.0000	0.0111
Total	0.0000	1.0000	0.0000	1.0000	0.0000	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.9798	---	0.9798
diM-Ether	11.73	---	11.73
Methanol	5.025	---	5.025

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Separator: V-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Waste Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Vapour Phase	Aqueous Phase
Vapour / Phase Fraction	0.0000	0.0000	1.0000
Temperature: (F)	344.8	344.8	344.8
Pressure: (psia)	128.4	128.4	128.4
Molar Flow (lbmole/hr)	582.2	0.0000	582.2
Mass Flow (lb/hr)	1.053e+004	0.0000	1.053e+004
Std Ideal Liq Vol Flow (barrel/day)	724.1	0.0000	724.1
Molar Enthalpy (Btu/lbmole)	-1.175e+005	-1.014e+005	-1.175e+005
Molar Entropy (Btu/lbmole-F)	9.146	29.14	9.146
Heat Flow (Btu/hr)	-6.840e+007	0.0000	-6.840e+007
Liq Vol Flow @Std Cond (barrel/day)	712.1	0.0000	712.1

PROPERTIES

	Overall	Vapour Phase	Aqueous Phase
Molecular Weight	18.09	18.37	18.09
Molar Density (lbmole/ft3)	3.031	1.487e-002	3.031
Mass Density (lb/ft3)	54.82	0.2732	54.82
Act. Volume Flow (barrel/day)	821.0	0.0000	821.0
Mass Enthalpy (Btu/lb)	-6496	-5518	-6496
Mass Entropy (Btu/lb-F)	0.5057	1.586	0.5057
Heat Capacity (Btu/lbmole-F)	19.01	12.28	19.01
Mass Heat Capacity (Btu/lb-F)	1.051	0.6684	1.051
LHV Molar Basis (Std) (Btu/lbmole)	1372	6893	1372
HHV Molar Basis (Std) (Btu/lbmole)	1.909e+004	2.497e+004	1.909e+004
HHV Mass Basis (Std) (Btu/lb)	1056	1359	1056
CO2 Loading	---	---	---
CO2 App ML Con (lbmole/ft3)	---	---	---
CO2 App WT Con (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	75.85	375.3	75.85
Phase Fraction [Vol. Basis]	---	---	1.000
Phase Fraction [Mass Basis]	0.0000	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	0.0000	1.000
Mass Exergy (Btu/lb)	49.92	---	---
Partial Pressure of CO2 (psia)	0.0000	---	---
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---	---
Avg. Liq. Density (lbmole/ft3)	3.437	3.355	3.437
Specific Heat (Btu/lbmole-F)	19.01	12.28	19.01
Std. Gas Flow (MMSCFD)	5.293	0.0000	5.293
Std. Ideal Liq. Mass Density (lb/ft3)	62.16	61.62	62.16
Act. Liq. Flow (USGPS)	0.3991	---	0.3991
Z Factor	---	1.000	4.906e-003
Watson K	10.63	10.63	10.63
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	---	---

Cp/(Cp - R) 1.117 1.193 1.117

Cp/Cv 1.213 1.193 1.213

Heat of Vap. (Btu/lbmole) 1.575e+004 --- ---

Kinematic Viscosity (cSt) 0.1026 2.591 0.1026

Liq. Mass Density (Std. Cond) (lb/ft3) 63.21 62.66 63.21

Liq. Vol. Flow (Std. Cond) (barrel/day) 712.1 0.0000 712.1

Liquid Fraction 1.000 0.0000 1.000

Molar Volume (ft3/lbmole) 0.3299 67.24 0.3299

Mass Heat of Vap. (Btu/lb) 870.9 --- ---

Phase Fraction [Molar Basis] 0.0000 0.0000 1.0000

Surface Tension (dyne/cm) 43.44 --- 43.44

Thermal Conductivity (Btu/hr-ft-F) 0.3918 1.826e-002 0.3918

Viscosity (cP) 9.011e-002 1.134e-002 9.011e-002

Cv (Semi-Ideal) (Btu/lbmole-F) 17.02 10.29 17.02

Mass Cv (Semi-Ideal) (Btu/lb-F) 0.9413 0.5602 0.9413

Cv (Btu/lbmole-F) 15.67 10.29 15.67

Mass Cv (Btu/lb-F) 0.8662 0.5602 0.8662

Cv (Ent. Method) (Btu/lbmole-F) --- --- ---

Mass Cv (Ent. Method) (Btu/lb-F) --- --- ---

Cp/Cv (Ent. Method) --- --- ---

Reid VP at 37.8 C (psia) 4.641 4.641 4.641

True VP at 37.8 C (psia) 1.003 1.199 1.003

Liq. Vol. Flow - Sum(Std. Cond) (barrel/day) 712.1 0.0000 712.1

Viscosity Index --- --- ---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	579.3219	0.9950	10436.5430	0.9911	716.0618	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	2.9112	0.0050	93.2800	0.0089	8.0269	0.0111
Total	582.2331	1.0000	10529.8230	1.0000	724.0886	1.0000

Vapour Phase Phase Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	0.0000	0.9749	0.0000	0.9562	0.0000	0.9456
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0251	0.0000	0.0438	0.0000	0.0544
Total	0.0000	1.0000	0.0000	1.0000	0.0000	1.0000

Aqueous Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	579.3219	0.9950	10436.5430	0.9911	716.0618	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Methanol	2.9112	0.0050	93.2800	0.0089	8.0269	0.0111
Total	582.2331	1.0000	10529.8230	1.0000	724.0886	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.9798	---	0.9798
diM-Ether	11.73	---	11.73
Methanol	5.025	---	5.025

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Cooler: E-102	Separator: V-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Material Stream: Cooled Waste Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

	Overall	Aqueous Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (F)	150.0	150.0
Pressure: (psia)	15.00	15.00
Molar Flow (lbmole/hr)	582.2	582.2
Mass Flow (lb/hr)	1.053e+004	1.053e+004
Std Ideal Liq Vol Flow (barrel/day)	724.1	724.1
Molar Enthalpy (Btu/lbmole)	-1.211e+005	-1.211e+005
Molar Entropy (Btu/lbmole-F)	4.016	4.016
Heat Flow (Btu/hr)	-7.049e+007	-7.049e+007
Liq Vol Flow @Std Cond (barrel/day)	712.1	712.1

PROPERTIES

	Overall	Aqueous Phase
Molecular Weight	18.09	18.09
Molar Density (lbmole/ft3)	3.361	3.361
Mass Density (lb/ft3)	60.79	60.79
Act. Volume Flow (barrel/day)	740.5	740.5
Mass Enthalpy (Btu/lb)	-6694	-6694
Mass Entropy (Btu/lb-F)	0.2221	0.2221
Heat Capacity (Btu/lbmole-F)	18.14	18.14
Mass Heat Capacity (Btu/lb-F)	1.003	1.003
LHV Molar Basis (Std) (Btu/lbmole)	1372	1372

HHV Molar Basis (Std) (Btu/lbmole)	1.909e+004	1.909e+004
HHV Mass Basis (Std) (Btu/lb)	1056	1056
CO2 Loading	---	---
CO2 App ML Con (lbmole/ft3)	---	---
CO2 App WT Con (lbmol/lb)	---	---
LHV Mass Basis (Std) (Btu/lb)	75.85	75.85
Phase Fraction [Vol. Basis]	0.0000	1.000
Phase Fraction [Mass Basis]	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	0.0000	1.000
Mass Exergy (Btu/lb)	4.108	---
Partial Pressure of CO2 (psia)	0.0000	---
Cost Based on Flow (Cost/s)	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---
Avg. Liq. Density (lbmole/ft3)	3.437	3.437
Specific Heat (Btu/lbmole-F)	18.14	18.14
Std. Gas Flow (MMSCFD)	5.293	5.293
Std. Ideal Liq. Mass Density (lb/ft3)	62.16	62.16
Act. Liq. Flow (USGPS)	0.3600	0.3600
Z Factor	6.821e-004	6.821e-004
Watson K	10.63	10.63
User Property	---	---
Partial Pressure of H2S (psia)	0.0000	---
Cp/(Cp - R)	1.123	1.123
Cp/Cv	1.173	1.173
Heat of Vap. (Btu/lbmole)	1.747e+004	---
Kinematic Viscosity (cSt)	0.4668	0.4668
Liq. Mass Density (Std. Cond) (lb/ft3)	63.21	63.21
Liq. Vol. Flow (Std. Cond) (barrel/day)	712.1	712.1
Liquid Fraction	1.000	1.000
Molar Volume (ft3/lbmole)	0.2975	0.2975
Mass Heat of Vap. (Btu/lb)	966.2	---
Phase Fraction [Molar Basis]	0.0000	1.0000
Surface Tension (dyne/cm)	64.75	64.75
Thermal Conductivity (Btu/hr-ft-F)	0.3786	0.3786
Viscosity (cP)	0.4545	0.4545
Cv (Semi-Ideal) (Btu/lbmole-F)	16.16	16.16
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.8934	0.8934
Cv (Btu/lbmole-F)	15.47	15.47
Mass Cv (Btu/lb-F)	0.8555	0.8555
Cv (Ent. Method) (Btu/lbmole-F)	80.64	80.64
Mass Cv (Ent. Method) (Btu/lb-F)	4.459	4.459
Cp/Cv (Ent. Method)	0.2250	0.2250
Reid VP at 37.8 C (psia)	4.641	4.641
True VP at 37.8 C (psia)	1.003	1.003
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	712.1	712.1
Viscosity Index	-9.985	---

COMPOSITION

Overall Phase Vapour Fraction 0.0000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	579.3219	0.9950	10436.5430	0.9911	716.0618	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	2.9112	0.0050	93.2800	0.0089	8.0269	0.0111
Total	582.2331	1.0000	10529.8230	1.0000	724.0886	1.0000

Aqueous Phase Phase Fraction 1.000

COMPONENTS	MOLAR FLOW (lbmole/hr)	MOLE FRACTION	MASS FLOW (lb/hr)	MASS FRACTION	LIQUID VOLUME FLOW (barrel/day)	LIQUID VOLUME FRACTION
H2O	579.3219	0.9950	10436.5430	0.9911	716.0618	0.9889
diM-Ether	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	2.9112	0.0050	93.2800	0.0089	8.0269	0.0111
Total	582.2331	1.0000	10529.8230	1.0000	724.0886	1.0000

K VALUE

COMPONENTS	MIXED	LIGHT	HEAVY
H2O	0.0000	---	0.0000
diM-Ether	0.0000	---	0.0000
Methanol	0.0000	---	0.0000

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Cooler: E-102	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Preheat Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation: E-100

Duty SP: 3.425e+007 Btu/hr Minimum Available Duty: 0.0000 Btu/hr Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Heater: E-100	Cooler: E-101	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Reboil 1 Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation: Reboiler @COL1

Duty SP: 8.445e+006 Btu/hr Minimum Available Duty: 0.0000 Btu/hr Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Distillation: T-100		

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Condense 1 Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation: Condenser @COL1

Duty SP: 7.292e+006 Btu/hr Minimum Available Duty: --- Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Distillation: T-100	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Reboil 2 Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation:

Duty SP: 1.336e+007 Btu/hr Minimum Available Duty: 0.0000 Btu/hr Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Distillation: T-101		

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Condense 2 Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation: Condenser @COL2

Duty SP: 1.352e+007 Btu/hr Minimum Available Duty: --- Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Distillation: T-101	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Feed Pump Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation: P-100

Duty SP: 1.352e+005 Btu/hr Minimum Available Duty: 0.0000 Btu/hr Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
Pump: P-100		

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

Energy Stream: Q Waste Cooler Fluid Package: Basis-1

Property Package: NRTL - Ideal

CONDITIONS

Duty Type: Direct Q Duty Calculation Operation: E-102

Duty SP: 2.085e+006 Btu/hr Minimum Available Duty: 0.0000 Btu/hr Maximum Available Duty: ---

COMPOSITION

(Not a material stream - No compositions exist)

UNIT OPERATIONS

FEED TO	PRODUCT FROM	LOGICAL CONNECTION
	Cooler: E-102	

UTILITIES

(No utilities reference this stream)

PROCESS UTILITY

CRV-100 (Conversion Reactor): Design, Reactions, Worksheet

Conversion Reactor: CRV-100

CONNECTIONS

Inlet Stream Connections

Stream Name	From Unit Operation
Preheated Reactor Feed	E-100 Heater

Outlet Stream Connections

Stream Name	To Unit Operation
Reactor Hot Gas Outlet	Cooler: E-101
Reactor Liquid Outlet	

Energy Stream Connections

Stream Name	From Unit Operation

PARAMETERS

Physical Parameters		Optional Heat Transfer	
Delta P	Vessel Volume	Duty	Energy Stream
0.0000 psi	---	0.0000 Btu/hr	

User Variables

REACTION DETAILS

Reaction: Rxn-1

Component	Mole Weight	Stoichiometric Coeff.
Methanol	32.04	-2.000
diM-Ether	46.07	1.000
H2O	18.02	1.000

REACTION RESULTS FOR : Set-1

Extents

Name	Rank	Specified % Conversion	Use Default	Actual % Conversion	Base Component	Reaction Extent
Rxn-1	0	70.64	Yes	70.64	Methanol	7.276e-002

Balance

Components	Total Inflow	Total Reaction	Total Outflow
H2O	4.116e-003	7.276e-002	7.687e-002
diM-Ether	5.259e-005	7.276e-002	7.281e-002
Methanol	0.2060	-0.1455	6.048e-002

CONDITIONS

Name	Preheated Reactor Feed	Reactor Liquid Outlet	Reactor Hot Gas Outlet
Vapour	1.0000	0.0000	1.0000
Temperature (F)	536.0000	717.5885	717.5885
Pressure (psia)	500.0000	500.0000	500.0000
Molar Flow (lbmole/hr)	1668.0006	0.0000	1668.0006
Mass Flow (lb/hr)	52993.5852	0.0000	52993.7581
Std Ideal Liq Vol Flow (barrel/day)	4550.2197	0.0000	4797.1286
Molar Enthalpy (Btu/lbmole)	-8.101e+004	-8.970e+004	-8.101e+004
Molar Entropy (Btu/lbmole-F)	42.56	43.68	45.95
Heat Flow (Btu/hr)	-1.3513e+08	0.0000e-01	-1.3513e+08

PROPERTIES

Name	Preheated Reactor Feed	Reactor Liquid Outlet	Reactor Hot Gas Outlet
Molecular Weight	31.77	26.70	31.77
Molar Density (lbmole/ft3)	4.679e-002	3.468	3.958e-002
Mass Density (lb/ft3)	1.487	92.59	1.257
Act. Volume Flow (barrel/day)	1.524e+005	0.0000	1.802e+005
Mass Enthalpy (Btu/lb)	-2550	-3360	-2550
Mass Entropy (Btu/lb-F)	1.340	1.636	1.446
Heat Capacity (Btu/lbmole-F)	15.04	16.75	17.38
Mass Heat Capacity (Btu/lb-F)	0.4733	0.6274	0.5471
LHV Molar Basis (Std) (Btu/lbmole)	2.690e+005	1.745e+005	2.768e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.040e+005	2.031e+005	3.117e+005
HHV Mass Basis (Std) (Btu/lb)	9567	7607	9812
CO2 Loading	---	---	---
CO2 Apparent Mole Conc. (lbmole/ft3)	---	---	---
CO2 Apparent Wt. Conc. (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	8468	6538	8713
Phase Fraction [Vol. Basis]	1.000	---	1.000
Phase Fraction [Mass Basis]	1.000	0.0000	1.000
Phase Fraction [Act. Vol. Basis]	1.000	---	1.000
Mass Exergy (Btu/lb)	171.6	159.4	244.8
Partial Pressure of CO2 (psia)	0.0000	0.0000	0.0000
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	594.1	---	702.4
Avg. Liq. Density (lbmole/ft3)	1.567	---	1.486
Specific Heat (Btu/lbmole-F)	15.04	16.75	17.38
Std. Gas Flow (MMSCFD)	15.16	0.0000	15.16
Std. Ideal Liq. Mass Density (lb/ft3)	49.78	50.31	47.22
Act. Liq. Flow (USGPS)	---	0.0000	0.0000
Z Factor	1.000	---	---
Watson K	10.63	11.18	11.19
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	0.0000	0.0000
Cp/(Cp - R)	1.152	1.135	1.129
Cp/Cv	1.152	1.135	1.129
Heat of Vap. (Btu/lbmole)	1.050e+004	1.342e+004	1.256e+004
Kinematic Viscosity (cSt)	0.5549	9.262e-002	0.9393

Liq. Mass Density (Std. Cond) (lb/ft3)	49.83	51.09	47.10
Liq. Vol. Flow (Std. Cond) (barrel/day)	4546	0.0000	4810
Liquid Fraction	0.0000	1.000	0.0000
Molar Volume (ft3/lbmole)	21.37	0.2883	25.27
Mass Heat of Vap. (Btu/lb)	330.6	502.8	395.5
Phase Fraction [Molar Basis]	1.0000	0.0000	1.0000
Surface Tension (dyne/cm)	---	---	---
Thermal Conductivity (Btu/hr-ft-F)	2.205e-002	7.587e-002	3.059e-002
Viscosity (cP)	1.322e-002	0.1374	1.892e-002
Cv (Semi-Ideal) (Btu/lbmole-F)	13.05	14.76	15.40
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.4108	0.5530	0.4846
Cv (Btu/lbmole-F)	13.05	14.76	15.40
Mass Cv (Btu/lb-F)	0.4108	0.5530	0.4846
Cv (Ent. Method) (Btu/lbmole-F)	---	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---	---
Cp/Cv (Ent. Method)	---	---	---
Reid VP at 37.8 C (psia)	4.669	62.96	65.93
True VP at 37.8 C (psia)	4.615	24.81	40.34
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	4546	0.0000	4810
Viscosity Index	---	---	---

T-100 (Distillation): Design, Profiles, Efficiencies, Solver, Internals, Rating, Worksheet, Performance, Flowsheet, Dynamics

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Distillation: T-100

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
Q Reboil 1	Reboiler	
Cooled Reactor Outlet	8__Main Tower	E-101 Cooler

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
Q Condense 1	Condenser	
DME Outlet	Condenser	
Methanol and Water Outlet	Reboiler	T-101 Distillation

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error
Reflux Ratio	0.2000	0.6034	2.017
Distillate Rate	578.0 lbmole/hr	578.0 lbmole/hr	5.497e-007
Reflux Rate	---	348.8 lbmole/hr	---
Btms Prod Rate	---	1090 lbmole/hr	---
Comp Recovery	0.9990	0.9996	2.906e-004

Comp Fraction	0.9995	0.9995	1.277e-004			
	Wt. Tol.	Abs. Tol.	Active	Estimate	Used	
Reflux Ratio	1.000e-002	1.000e-002	Off	On	Off	
Distillate Rate	1.000e-002	2.205 lbmole/hr	On	On	On	
Reflux Rate	1.000e-002	2.205 lbmole/hr	Off	On	Off	
Btms Prod Rate	1.000e-002	2.205 lbmole/hr	Off	Off	Off	
Comp Recovery	1.000e-002	1.000e-003	Off	On	Off	
Comp Fraction	1.000e-002	1.000e-003	On	On	On	

SPECS

Column Specification Parameters

Reflux Ratio

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Condenser Flow Basis: Molar Liquid Spec: ---

Distillate Rate

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stream: DME Outlet @COL1 Flow Basis: Molar

Reflux Rate

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Condenser Flow Basis: Molar Liquid Spec: ---

Btms Prod Rate

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stream: Methanol and Water Outlet @COL1 Flow Basis: Molar

Comp Recovery

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Draw: DME Outlet @COL1 Flow Basis: Molar

Components: diM-Ether

Comp Fraction

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Condenser Flow Basis: Mass Fraction Phase: Liquid

Components: diM-Ether

SUBCOOLING

	Condenser
Degrees of Subcooling	---
Subcool to	---

User Variables

PROFILES

General Parameters

Sub-Flow Sheet: T-100 (COL1) Number of Stages: 23

Profile Estimates

	Temperature (F)	Net Liquid (lbmole/hr)	Net Vapour (lbmole/hr)
Condenser	117.3	348.8	1.374e-006
1__Main Tower	117.6	345.5	926.8
2__Main Tower	118.8	335.4	923.5
3__Main Tower	122.5	306.3	913.4
4__Main Tower	133.5	248.8	884.4
5__Main Tower	156.7	195.7	826.9
6__Main Tower	179.6	170.6	773.8
7__Main Tower	192.3	155.6	748.6
8__Main Tower	202.1	1817	733.6
9__Main Tower	204.6	1809	726.6
10__Main Tower	208.9	1797	719.0
11__Main Tower	216.3	1780	707.1
12__Main Tower	227.4	1761	690.3
13__Main Tower	241.9	1744	670.8
14__Main Tower	257.7	1736	654.2
15__Main Tower	273.1	1709	645.9
16__Main Tower	282.7	1707	619.2
17__Main Tower	289.4	1708	616.8
18__Main Tower	293.6	1710	618.2
19__Main Tower	296.1	1711	619.8
20__Main Tower	297.6	1711	620.9
21__Main Tower	298.5	1711	621.4
22__Main Tower	299.3	1709	621.1
23__Main Tower	300.5	1702	619.1
Reboiler	305.2	1090	612.3

EFFICIENCIES

Stage Efficiencies

Stages	Overall	H2O	diM-Ether	Methanol
Condenser	1.000	1.000	1.000	1.000
1__Main Tower	0.7700	0.7700	0.7700	0.7700
2__Main Tower	0.7700	0.7700	0.7700	0.7700
3__Main Tower	0.7700	0.7700	0.7700	0.7700
4__Main Tower	0.7700	0.7700	0.7700	0.7700
5__Main Tower	0.7700	0.7700	0.7700	0.7700
6__Main Tower	0.7700	0.7700	0.7700	0.7700
7__Main Tower	0.7700	0.7700	0.7700	0.7700
8__Main Tower	0.7700	0.7700	0.7700	0.7700
9__Main Tower	0.7700	0.7700	0.7700	0.7700
10__Main Tower	0.7700	0.7700	0.7700	0.7700
11__Main Tower	0.7700	0.7700	0.7700	0.7700
12__Main Tower	0.7700	0.7700	0.7700	0.7700
13__Main Tower	0.7700	0.7700	0.7700	0.7700
14__Main Tower	0.7700	0.7700	0.7700	0.7700
15__Main Tower	0.7700	0.7700	0.7700	0.7700

16__Main Tower	0.7700	0.7700	0.7700	0.7700
17__Main Tower	0.7700	0.7700	0.7700	0.7700
18__Main Tower	0.7700	0.7700	0.7700	0.7700
19__Main Tower	0.7700	0.7700	0.7700	0.7700
20__Main Tower	0.7700	0.7700	0.7700	0.7700
21__Main Tower	0.7700	0.7700	0.7700	0.7700
22__Main Tower	0.7700	0.7700	0.7700	0.7700
23__Main Tower	0.7700	0.7700	0.7700	0.7700
Reboiler	1.000	1.000	1.000	1.000

SOLVER

Column Solving Algorithm: HYSIM Inside-Out

Solving Options | Acceleration Parameters

Maximum Iterations: 10000 | Accelerate K Value & H Model Parameters: Off

Equilibrium Error Tolerance: 1.000e-05

Heat/Spec Error Tolerance: 5.000e-004

Save Solutions as Initial Estimate: On

Super Critical Handling Model: Simple K

Trace Level: Low

Init from Ideal K's: Off | Damping Parameters

Initial Estimate Generator Parameters | Azeotrope Check: Off

Iterative IEG (Good for Chemicals): Off | Fixed Damping Factor: 1

ACTIVE INTERNAL OPTION: Internals-1@Main Tower@COL1

Name	Start Stage	End Stage	Mode	Internals	Tray / Packing Type	Number of Passes	Packing Vendor	Packing Material	Packing Dimension	Tray Spacing / Section Packed Height (ft)	Diameter (ft)
CS-1	1__Main Tower	23__Main Tower	Interactive Sizing	Trayed	Sieve	1	---	---	---	2.000	3.000

SETUP

Section Name CS-1

Section Start 1__Main Tower

Section End 23__Main Tower

Internals Trayed

Internals Type Sieve

Diameter (ft) 3.000

Tray Spacing / Section Packed Height (ft) 2.000

Number Of Passes 1

Maximum Acceptable Pressure Drop (psi) 0.3626

Maximum Percent Downcomer Backup 100.00 %

Maximum Percent Jet Flood 100.00 %

Percent Jet Flood For Design 80.00 %

Maximum Percent Liquid Entrainment 10.00 %

Minimum Weir Loading (USGPM/ft) 6.000

Maximum Weir Loading (USGPM/ft) 157.5

Minimum Downcomer Area / Total Tray Area 0.1000

Override Downcomer Froth Density No

Froth Density ---

Weep Method Hsieh

Default Jet Flood Calculation Method GLITSCH6

Maximum Downcomer Loading Method Glitsch

% Approach to Maximum Capacity ---

Design Capacity Factor ---

Capacity Factor at Flooding ---

System Foaming Factor 1.000

Aeration Factor Multipler 1.000

Minimum Liquid Flow Rate ---

Pressure Drop at Flood per Unit Packed Height ---

Allowable Pressure Drop per Unit Packed Height ---

Minimum Pressure Drop per Unit Packed Height ---

Number of Curves ---

Warning Status (% to Limit) 10.00 %

Pressure Drop Calculation Method ---

Mode Interactive Sizing

Status Needs Calculating

GEOMETRY DETAILS

Common Geometry CS-1

Section Start 1__Main Tower

Section End 23__Main Tower

Internals Sieve

Section Diameter (ft) 3.000

Foaming Factor 1.000

Over-Design Factor 1.000

Common Tray Geometry CS-1

Number of Passes 1

Tray Spacing (ft) 2.000

Picket Fence Weirs No

Swept Back Weirs Yes

Active Area Under Downcomer No

Deck Thickness 10 Gauge

Deck Thickness Value (in) 0.1340

Balance Downcomers Based On Maximum Downcomer Loading

Weir Modifications Swept-Back

Net Area (ft2) 6.414

Cross-Sectional Area (ft2) 7.069

Active Area (ft2) 5.760

Downcomer Geometry CS-1

Side Weir Height (in) 2.756

Weir Length (ft) ---

Downcomer Clearance (in) 2.756

Downcomer Width - Top (in) 5.340

Downcomer Width - Bottom (in) 5.340

Downcomer Loading Top (USGPM/ft2) 85.25

Weir Loading (USGPM/ft) 82.59

Downcomer Area - Top (ft2) 0.6541

Downcomer Area - Bottom (ft2) 0.6541

Picketing Fraction ---

Center Weir Height (in) ---

Weir Length (ft) ---

Downcomer Clearance (in) ---

Downcomer Width - Top (in) ---

Downcomer Width - Bottom (in) ---

Downcomer Loading Top (USGPM/ft2) ---

Weir Loading (USGPM/ft) ---

Downcomer Area - Top (ft2) ---

Downcomer Area - Bottom (ft2) ---

Picketing Fraction ---

Off Center Weir Height (in) ---

Inside Weir Length (ft) ---

Outside Weir Length (ft) ---

Downcomer Clearance (in) ---

Downcomer Width - Top (in) ---

Downcomer Width - Bottom (in) ---

Downcomer Loading Top (USGPM/ft2) ---

Maximum Outside Weir Loading (USGPM/ft) ---

Maximum Inside Weir Loading (USGPM/ft) ---

Downcomer Area - Top (ft2) ---

Downcomer Area - Bottom (ft2) ---

Inside Picketing Fraction ---

Outside Picketing Fraction ---

Off-Center Downcomer Location (ft) ---

Swept Back Weir Geometry CS-1

Compatibility KG Tower

A ---

B/Parallel Chord Segment ---

S/Swept-Back Weir 0.0000

Swept-Back Weir Chord ---

Angled Chord Segment ---

Tray With Maximum Weir Loading 8

Maximum Weir Loading (USGPM/ft) 82.59

Maximum Allowable Weir Loading in Section (USGPM/ft) 157.5

Actual Side Weir Length (ft) 2.133

Effective Side Weir Length (ft) 2.133

Lost Area (%) 0.00

Sieve Geometry CS-1

Hole Diameter (in) 0.5000

Number of Holes 422

Hole Area to Active Area 0.1000

Bubble Cap Geometry CS-1

Cap Diameter ---

Skirt Height ---

Number of Caps ---

Number of Caps Per Active Area ---

Valve Geometry CS-1

Tray Type ---

Valve Type ---

Valve Material ---

Leg Length ---

Valve Thickness ---

Number of Valves ---

Number of Valves per Active Area ---

Packing Geometry CS-1

HETP (ft) ---

Section Packed Height (ft) ---

Packing Type ---

Packing Vendor ---

Packing Material ---

Packing Dimension ---

Packing Factor (ft2/ft3) ---

Packing Surface Area (ft2/ft3) ---

1st Stichlmair Constant ---

2nd Stichlmair Constant ---

3rd Stichlmair Constant ---

Void Fraction ---

RESULTS SUMMARY

Section Name CS-1

Section Start 1__Main Tower

Section End 23__Main Tower

Internals Trayed

Diameter (ft) 3.000

Number of Passes 1

Tray Spacing / Section Packed Height (ft) 2.000

Total Height (ft) 46.00

Total Pressure Drop (psi) 61.56

Total Pressure Drop (Head Loss) (ft) 93.69

Trays With Weeping None

Maximum Percent Jet Flood (%) 78.49

Tray With Maximum Jet Flood 1__Main Tower

Maximum Percent Downcomer Backup (%) 37.77

Tray With Maximum Downcomer Backup 8__Main Tower

Maximum Downcomer Loading (USGPM/ft2) 269.2

Tray With Maximum Downcomer Loading 8__Main Tower

Maximum Downcomer Loading Location Side

Maximum Weir Loading (USGPM/ft) 82.59

Tray With Maximum Weir Loading 8__Main Tower

Maximum Weir Loading Location Side

Maximum Aerated Height Over Weir (in) 7.401

Tray With Maximum Aerated Height Over Weir 8__Main Tower

Maximum % Approach To System Limit (%) 64.45

Tray With Maximum % Approach To System Limit 1__Main Tower

Maximum Cs Based On Bubbling Area (%) 0.3273

Tray With Maximum Cs Based On Bubbling Area 1__Main Tower

Maximum % Capacity (Constant L/V) 78.49

Maximum Capacity Factor ---

Section Pressure Drop (psi) 61.56

Average Pressure Drop Per Height (inH2O/ft) ---

Average Pressure Drop Per Height (Frictional) (inH2O/ft) ---

Maximum Stage Liquid Holdup (ft3) ---

Maximum Liquid Superficial Velocity (ft/s) ---

Surface Area (ft2/ft3) ---

Void Fraction ---

1st Stichlmair Constant ---

2nd Stichlmair Constant ---

3rd Stichlmair Constant ---

STAGE BY STAGE RESULTS: CS-1

State Conditions

Stages	Liquid Temperature (F)	Vapor Temperature (F)	Liquid Mass Flow (lb/hr)	Vapor Mass Flow (lb/hr)	Liquid Volume Flow (USGPS)	Vapor Volume Flow (USGPS)
1__Main Tower	117.6	118.8	1.589e+004	4.252e+004	0.9294	77.44
2__Main Tower	118.8	122.5	1.536e+004	4.198e+004	0.8965	77.02
3__Main Tower	122.5	133.5	1.384e+004	4.046e+004	0.8023	75.94
4__Main Tower	133.5	156.7	1.078e+004	3.741e+004	0.6122	73.73
5__Main Tower	156.7	179.6	7820	3.444e+004	0.4275	71.51
6__Main Tower	179.6	192.3	6326	3.295e+004	0.3361	70.51
7__Main Tower	192.3	202.1	5465	3.209e+004	0.2854	70.10
8__Main Tower	202.1	204.6	5.801e+004	3.164e+004	2.935	69.63
9__Main Tower	204.6	208.9	5.741e+004	3.104e+004	2.898	69.31
10__Main Tower	208.9	216.3	5.642e+004	3.005e+004	2.837	68.87
11__Main Tower	216.3	227.4	5.491e+004	2.854e+004	2.743	68.29
12__Main Tower	227.4	241.9	5.291e+004	2.654e+004	2.618	67.72
13__Main Tower	241.9	257.7	5.072e+004	2.436e+004	2.479	67.49
14__Main Tower	257.7	273.1	4.880e+004	2.243e+004	2.354	68.01
15__Main Tower	273.1	282.7	4.658e+004	2.021e+004	2.221	66.01
16__Main Tower	282.7	289.4	4.558e+004	1.921e+004	2.154	66.31
17__Main Tower	289.4	293.6	4.501e+004	1.864e+004	2.115	66.78
18__Main Tower	293.6	296.1	4.470e+004	1.833e+004	2.094	67.13
19__Main Tower	296.1	297.6	4.451e+004	1.814e+004	2.081	67.34
20__Main Tower	297.6	298.5	4.439e+004	1.802e+004	2.073	67.43
21__Main Tower	298.5	299.3	4.426e+004	1.789e+004	2.064	67.42
22__Main Tower	299.3	300.5	4.403e+004	1.766e+004	2.049	67.26
23__Main Tower	300.5	305.2	4.346e+004	1.709e+004	2.014	66.94

Physical Conditions

Stages Liquid Molecular Weight Vapor Molecular Weight Liquid Mass Density Vapor Mass Density Liquid Viscosity Vapor Viscosity Surface Tension

(lb/ft3) (lb/ft3) (cP) (cP) (dyne/cm)

1__Main Tower	46.00	46.04	35.53	1.141	7.306e-002	9.818e-003	8.762
2__Main Tower	45.80	45.96	35.60	1.133	7.433e-002	9.859e-003	8.925
3__Main Tower	45.18	45.75	35.85	1.107	7.843e-002	9.984e-003	9.455
4__Main Tower	43.33	45.24	36.60	1.054	9.118e-002	1.023e-002	11.04
5__Main Tower	39.96	44.52	38.01	1.001	0.1155	1.043e-002	13.90
6__Main Tower	37.09	44.01	39.12	0.9710	0.1328	1.056e-002	16.64
7__Main Tower	35.12	43.74	39.78	0.9512	0.1380	1.076e-002	19.85
8__Main Tower	31.93	43.54	41.06	0.9441	0.1373	1.076e-002	27.65
9__Main Tower	31.73	43.17	41.16	0.9305	0.1387	1.075e-002	27.75
10__Main Tower	31.39	42.49	41.32	0.9066	0.1419	1.073e-002	27.90
11__Main Tower	30.84	41.34	41.59	0.8684	0.1431	1.067e-002	28.12
12__Main Tower	30.05	39.56	41.99	0.8144	0.1445	1.054e-002	28.36
13__Main Tower	29.08	37.23	42.52	0.7499	0.1454	1.032e-002	28.52
14__Main Tower	28.11	34.72	43.08	0.6852	0.1452	1.007e-002	28.51
15__Main Tower	27.26	32.65	43.57	0.6363	0.1445	9.808e-003	28.38
16__Main Tower	26.70	31.14	43.97	0.6019	0.1460	9.624e-003	28.23
17__Main Tower	26.35	30.16	44.22	0.5801	0.1467	9.506e-003	28.04
18__Main Tower	26.14	29.57	44.36	0.5672	0.1470	9.437e-003	27.90
19__Main Tower	26.02	29.22	44.45	0.5598	0.1473	9.400e-003	27.82
20__Main Tower	25.94	29.00	44.51	0.5553	0.1474	9.384e-003	27.79
21__Main Tower	25.87	28.81	44.56	0.5515	0.1474	9.389e-003	27.83
22__Main Tower	25.76	28.53	44.64	0.5456	0.1473	9.428e-003	28.01
23__Main Tower	25.53	27.91	44.84	0.5304	0.1468	9.573e-003	28.52

Hydraulic Results

Stages	Percent Jet Flood	Dry Pressure Drop	Total Pressure Drop	Dry Pressure Drop (Head Loss)	Total Pressure Drop (Head Loss)
	(%)	(inH2O(60F))	(inH2O(60F))	(in)	(in)
1__Main Tower	78.49	2.032	3.075	3.567	5.398
2__Main Tower	77.56	1.996	3.039	3.496	5.323
3__Main Tower	74.94	1.896	2.938	3.299	5.111
4__Main Tower	69.42	1.702	2.747	2.900	4.681
5__Main Tower	63.55	1.520	2.588	2.494	4.246
6__Main Tower	60.40	1.434	2.524	2.286	4.025
7__Main Tower	58.67	1.388	2.496	2.176	3.913
8__Main Tower	69.26	1.360	2.987	2.065	4.537
9__Main Tower	68.32	1.328	2.959	2.012	4.483
10__Main Tower	66.81	1.277	2.914	1.928	4.398
11__Main Tower	64.51	1.203	2.847	1.803	4.270
12__Main Tower	61.49	1.109	2.765	1.647	4.107
13__Main Tower	58.24	1.014	2.683	1.488	3.935
14__Main Tower	55.49	0.9414	2.619	1.363	3.791
15__Main Tower	51.65	0.8234	2.534	1.179	3.627
16__Main Tower	50.12	0.7860	2.507	1.115	3.555
17__Main Tower	49.34	0.7684	2.493	1.084	3.516
18__Main Tower	48.91	0.7593	2.485	1.067	3.494
19__Main Tower	48.67	0.7541	2.481	1.058	3.481
20__Main Tower	48.49	0.7500	2.478	1.051	3.473
21__Main Tower	48.28	0.7445	2.475	1.042	3.464
22__Main Tower	47.86	0.7331	2.469	1.024	3.449

23__Main Tower	46.87	0.7060	2.455	0.9819	3.414

Stages	Downcomer Backup (Aerated)	Downcomer Backup (Unaerated)	Percent Downcomer Backup (Aerated)	Percent Downcomer Backup (Unaerated)
	(ft)	(ft)	(%)	(%)
1__Main Tower	0.7969	0.4684	35.74	21.01
2__Main Tower	0.7866	0.4626	35.28	20.75
3__Main Tower	0.7579	0.4465	33.99	20.03
4__Main Tower	0.7002	0.4146	31.40	18.59
5__Main Tower	0.6432	0.3836	28.85	17.20
6__Main Tower	0.6155	0.3686	27.60	16.53
7__Main Tower	0.6020	0.3613	27.00	16.21
8__Main Tower	0.8421	0.5071	37.77	22.74
9__Main Tower	0.8357	0.5033	37.48	22.57
10__Main Tower	0.8254	0.4974	37.02	22.31
11__Main Tower	0.8105	0.4887	36.35	21.92
12__Main Tower	0.7920	0.4779	35.52	21.44
13__Main Tower	0.7733	0.4671	34.68	20.95
14__Main Tower	0.7580	0.4583	34.00	20.55
15__Main Tower	0.7429	0.4494	33.32	20.16
16__Main Tower	0.7360	0.4455	33.01	19.98
17__Main Tower	0.7321	0.4433	32.84	19.88
18__Main Tower	0.7300	0.4420	32.74	19.83
19__Main Tower	0.7287	0.4413	32.68	19.79
20__Main Tower	0.7279	0.4409	32.65	19.77
21__Main Tower	0.7272	0.4405	32.62	19.75
22__Main Tower	0.7261	0.4398	32.56	19.72
23__Main Tower	0.7235	0.4383	32.45	19.66

Stages	Mass Rate / Column Area	Volume Rate / Column Area	Fs (Net Area)	Fs (Bubble Area)	Cs (Net Area)
	(lb/s-ft2)	(USGPM/ft2)	(ft/(s/sqrt(lb/ft3)))	(ft/(s/sqrt(lb/ft3)))	(ft/s)
1__Main Tower	0.6245	7.889	1.724	1.920	0.2940
2__Main Tower	0.6036	7.610	1.708	1.902	0.2910
3__Main Tower	0.5439	6.810	1.665	1.854	0.2825
4__Main Tower	0.4238	5.196	1.578	1.757	0.2646
5__Main Tower	0.3073	3.629	1.491	1.660	0.2451
6__Main Tower	0.2486	2.853	1.448	1.613	0.2345
7__Main Tower	0.2148	2.423	1.425	1.587	0.2286
8__Main Tower	2.280	24.92	1.410	1.570	0.2226
9__Main Tower	2.256	24.60	1.393	1.552	0.2197
10__Main Tower	2.217	24.08	1.367	1.522	0.2150
11__Main Tower	2.158	23.29	1.326	1.477	0.2078
12__Main Tower	2.079	22.22	1.274	1.418	0.1985
13__Main Tower	1.993	21.04	1.218	1.356	0.1885
14__Main Tower	1.918	19.98	1.173	1.307	0.1802
15__Main Tower	1.831	18.86	1.097	1.222	0.1675
16__Main Tower	1.791	18.28	1.072	1.194	0.1628
17__Main Tower	1.769	17.96	1.060	1.180	0.1605
18__Main Tower	1.756	17.77	1.054	1.173	0.1592
19__Main Tower	1.749	17.66	1.050	1.169	0.1585

20__Main Tower	1.744	17.59	1.047	1.166	0.1580
21__Main Tower	1.739	17.52	1.043	1.162	0.1573
22__Main Tower	1.730	17.40	1.035	1.153	0.1559
23__Main Tower	1.708	17.09	1.016	1.131	0.1526

Stages	Cs (Bubble Area) (ft/s)	Approach to System Limit (%)	Height Over Weir (Aerated) (ft)	Height Over Weir (Unaerated) (ft)
1__Main Tower	0.3273	64.45	0.2838	5.445e-002
2__Main Tower	0.3240	63.49	0.2651	5.112e-002
3__Main Tower	0.3146	60.74	0.2280	4.460e-002
4__Main Tower	0.2947	54.74	0.1725	3.481e-002
5__Main Tower	0.2729	48.05	0.1260	2.633e-002
6__Main Tower	0.2611	44.12	0.1037	2.209e-002
7__Main Tower	0.2546	41.26	9.133e-002	1.966e-002
8__Main Tower	0.2479	37.69	0.6167	0.1337
9__Main Tower	0.2446	37.14	0.6083	0.1329
10__Main Tower	0.2394	36.27	0.5945	0.1316
11__Main Tower	0.2314	34.96	0.5734	0.1296
12__Main Tower	0.2210	33.27	0.5455	0.1268
13__Main Tower	0.2099	31.50	0.5150	0.1236
14__Main Tower	0.2007	30.08	0.4887	0.1204
15__Main Tower	0.1865	27.96	0.4552	0.1178
16__Main Tower	0.1813	27.21	0.4409	0.1160
17__Main Tower	0.1787	26.86	0.4332	0.1150
18__Main Tower	0.1773	26.68	0.4290	0.1143
19__Main Tower	0.1765	26.57	0.4265	0.1140
20__Main Tower	0.1759	26.49	0.4248	0.1137
21__Main Tower	0.1752	26.36	0.4229	0.1135
22__Main Tower	0.1736	26.09	0.4193	0.1132
23__Main Tower	0.1700	25.42	0.4106	0.1124

Side Downcomer Results

Stages	Volume (ft3)	Residence Time (seconds)	Velocity From Top (ft/s)	Velocity from Bottom (ft/s)	Exit Velocity (ft/s)
1__Main Tower	0.3064	2.466	0.1899	0.1899	0.2537
2__Main Tower	0.3026	2.525	0.1832	0.1832	0.2447
3__Main Tower	0.2921	2.724	0.1640	0.1640	0.2190
4__Main Tower	0.2712	3.314	0.1251	0.1251	0.1671
5__Main Tower	0.2509	4.391	8.737e-002	8.737e-002	0.1167
6__Main Tower	0.2411	5.368	6.868e-002	6.868e-002	9.173e-002
7__Main Tower	0.2364	6.194	5.833e-002	5.833e-002	7.791e-002
8__Main Tower	0.3317	0.8453	0.5999	0.5999	0.8012
9__Main Tower	0.3292	0.8498	0.5923	0.5923	0.7911
10__Main Tower	0.3253	0.8578	0.5798	0.5798	0.7744
11__Main Tower	0.3197	0.8717	0.5606	0.5606	0.7488
12__Main Tower	0.3126	0.8933	0.5350	0.5350	0.7146
13__Main Tower	0.3056	0.9220	0.5066	0.5066	0.6766
14__Main Tower	0.2998	0.9528	0.4810	0.4810	0.6424
15__Main Tower	0.2940	0.9900	0.4540	0.4540	0.6063
16__Main Tower	0.2914	1.012	0.4401	0.4401	0.5879

17__Main Tower	0.2900	1.025	0.4323	0.4323	0.5774
18__Main Tower	0.2892	1.033	0.4278	0.4278	0.5714
19__Main Tower	0.2887	1.038	0.4253	0.4253	0.5680
20__Main Tower	0.2884	1.041	0.4235	0.4235	0.5657
21__Main Tower	0.2881	1.044	0.4218	0.4218	0.5634
22__Main Tower	0.2877	1.050	0.4188	0.4188	0.5594
23__Main Tower	0.2867	1.065	0.4115	0.4115	0.5497

RATING

Tray Sections

Tray Section	Main Tower @COL1
Tray Diameter (ft)	4.921
Weir Height (ft)	0.1640
Weir Length (ft)	3.937
Tray Space (ft)	1.804
Tray Volume (ft3)	34.32
Disable Heat Loss Calculations	No
Heat Model	None
Rating Calculations	No
Tray Hold Up (ft3)	3.120

Vessels

Vessel	Reboiler @COL1	Condenser @COL1
Diameter (ft)	3.914	3.914
Length (ft)	5.871	5.871
Volume (ft3)	70.63	70.63
Orientation	Horizontal	Horizontal
Vessel has a Boot	No	No
Boot Diameter (ft)	---	---
Boot Length (ft)	---	---
Hold Up (ft3)	35.31	35.31

Other Equipment In Column Flowsheet

Pressure Profile

	Pressure (psia)	Pressure Drop (psi)
Condenser	153.7 psia	0.0000 psi
1__Main Tower	153.7 psia	0.1027 psi
2__Main Tower	153.8 psia	0.1027 psi
3__Main Tower	153.9 psia	0.1027 psi
4__Main Tower	154.0 psia	0.1027 psi
5__Main Tower	154.2 psia	0.1027 psi
6__Main Tower	154.3 psia	0.1027 psi
7__Main Tower	154.4 psia	0.1027 psi
8__Main Tower	154.5 psia	0.1027 psi
9__Main Tower	154.6 psia	0.1027 psi

10__Main Tower	154.7 psia	0.1027 psi
11__Main Tower	154.8 psia	0.1027 psi
12__Main Tower	154.9 psia	0.1027 psi
13__Main Tower	155.0 psia	0.1027 psi
14__Main Tower	155.1 psia	0.1027 psi
15__Main Tower	155.2 psia	0.1027 psi
16__Main Tower	155.3 psia	0.1027 psi
17__Main Tower	155.4 psia	0.1027 psi
18__Main Tower	155.5 psia	0.1027 psi
19__Main Tower	155.6 psia	0.1027 psi
20__Main Tower	155.7 psia	0.1027 psi
21__Main Tower	155.8 psia	0.1027 psi
22__Main Tower	155.9 psia	0.1027 psi
23__Main Tower	156.0 psia	---
Reboiler	156.0 psia	0.0000 psi

Pressure Solving Options

Pressure Tolerance 1.000e-004 Pressure Drop Tolerance 1.000e-004

Damping Factor 1.000 Max Press Iterations 100

CONDITIONS

Name	Cooled Reactor Outlet	Methanol and Water Outlet	DME Outlet	Q Reboil 1	Q Condense 1
Vapour	0.0000	0.0000	0.0000	---	---
Temperature (F)	201.9609	305.2081	117.2574	---	---
Pressure (psia)	500.0000	156.0000	153.7400	---	---
Molar Flow (lbmole/hr)	1668.0006	1089.9542	578.0464	---	---
Mass Flow (lb/hr)	52993.7581	26369.6753	26624.0827	---	---
Std Ideal Liq Vol Flow (barrel/day)	4797.1286	2077.6358	2719.4928	---	---
Molar Enthalpy (Btu/lbmole)	-1.016e+005	-1.085e+005	-8.641e+004	---	---
Molar Entropy (Btu/lbmole-F)	28.90	21.07	38.59	---	---
Heat Flow (Btu/hr)	-1.6939e+08	-1.1829e+08	-4.9948e+07	8.4452e+06	7.2925e+06

PROPERTIES

Name	Cooled Reactor Outlet	Methanol and Water Outlet	DME Outlet
Molecular Weight	31.77	24.19	46.06
Molar Density (lbmole/ft3)	1.304	1.907	0.7709
Mass Density (lb/ft3)	41.43	46.14	35.51
Act. Volume Flow (barrel/day)	5467	2443	3205
Mass Enthalpy (Btu/lb)	-3196	-4486	-1876
Mass Entropy (Btu/lb-F)	0.9095	0.8709	0.8378
Heat Capacity (Btu/lbmole-F)	28.21	25.61	28.61
Mass Heat Capacity (Btu/lb-F)	0.8880	1.059	0.6212
LHV Molar Basis (Std) (Btu/lbmole)	2.768e+005	1.208e+005	5.709e+005
HHV Molar Basis (Std) (Btu/lbmole)	3.117e+005	1.462e+005	6.238e+005
HHV Mass Basis (Std) (Btu/lb)	9812	6044	1.354e+004
CO2 Loading	---	---	---
CO2 Apparent Mole Conc. (lbmole/ft3)	---	---	---
CO2 Apparent Wt. Conc. (lbmol/lb)	---	---	---

LHV Mass Basis (Std) (Btu/lb)	8713	4995	1.240e+004
Phase Fraction [Vol. Basis]	0.0000	0.0000	0.0000
Phase Fraction [Mass Basis]	0.0000	0.0000	0.0000
Phase Fraction [Act. Vol. Basis]	0.0000	0.0000	0.0000
Mass Exergy (Btu/lb)	-113.5	-87.38	-65.72
Partial Pressure of CO2 (psia)	0.0000	0.0000	0.0000
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	---	---	---
Avg. Liq. Density (lbmole/ft3)	1.486	2.243	0.9086
Specific Heat (Btu/lbmole-F)	28.21	25.61	28.61
Std. Gas Flow (MMSCFD)	15.16	9.908	5.254
Std. Ideal Liq. Mass Density (lb/ft3)	47.22	54.25	41.85
Act. Liq. Flow (USGPS)	2.658	1.188	1.558
Z Factor	5.400e-002	9.966e-003	3.221e-002
Watson K	11.19	10.63	11.38
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	0.0000	0.0000
Cp/(Cp - R)	1.076	1.084	1.075
Cp/Cv	1.248	1.233	1.075
Heat of Vap. (Btu/lbmole)	1.256e+004	1.474e+004	7879
Kinematic Viscosity (cSt)	0.2062	0.1957	0.1278
Liq. Mass Density (Std. Cond) (lb/ft3)	47.10	54.76	38.58
Liq. Vol. Flow (Std. Cond) (barrel/day)	4810	2058	2950
Liquid Fraction	1.000	1.000	1.000
Molar Volume (ft3/lbmole)	0.7668	0.5244	1.297
Mass Heat of Vap. (Btu/lb)	395.5	609.1	171.1
Phase Fraction [Molar Basis]	0.0000	0.0000	0.0000
Surface Tension (dyne/cm)	28.24	31.66	8.712
Thermal Conductivity (Btu/hr-ft-F)	0.1408	0.2087	6.851e-002
Viscosity (cP)	0.1368	0.1447	7.266e-002
Cv (Semi-Ideal) (Btu/lbmole-F)	26.23	23.62	26.62
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.8255	0.9765	0.5780
Cv (Btu/lbmole-F)	22.60	20.78	26.62
Mass Cv (Btu/lb-F)	0.7114	0.8588	0.5780
Cv (Ent. Method) (Btu/lbmole-F)	124.2	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	3.908	---	---
Cp/Cv (Ent. Method)	0.2272	---	---
Reid VP at 37.8 C (psia)	65.93	4.697	120.5
True VP at 37.8 C (psia)	40.34	3.173	120.5
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	4810	2058	2950
Viscosity Index	-63.00	---	---

SUMMARY

Flow Basis: Molar The composition option is selected

Feed Composition

Cooled Reactor Outlet

Flow Rate (lbmole/hr) 1.668001e+03

H2O	0.3658
diM-Ether	0.3464
Methanol	0.2878

Flow Basis: Molar The composition option is selected

Feed Flows

	Cooled Reactor Outlet
Flow Rate (lbmole/hr)	1.668001e+03

H2O (lbmole/hr)	610.1179
diM-Ether (lbmole/hr)	577.8707
Methanol (lbmole/hr)	480.0121

Products

Flow Basis: Molar The composition option is selected

Product Compositions

	DME Outlet	Methanol and Water Outlet
Flow Rate (lbmole/hr)	578.0464	1.089954e+03
	---	---
H2O	0.0000	0.5598
diM-Ether	0.9993	0.0002
Methanol	0.0007	0.4400

Flow Basis: Molar The composition option is selected

Product Flows

	DME Outlet	Methanol and Water Outlet
Flow Rate (lbmole/hr)	578.0464	1.089954e+03
	---	---
H2O (lbmole/hr)	0.0055	610.1124
diM-Ether (lbmole/hr)	577.6286	0.2421
Methanol (lbmole/hr)	0.4123	479.5998

Flow Basis: Molar The composition option is selected

Product Recoveries

	DME Outlet	Methanol and Water Outlet
Flow Rate (lbmole/hr)	578.0464	1.089954e+03
	---	---
H2O (%)	0.0009	99.9991
diM-Ether (%)	99.9581	0.0419
Methanol (%)	0.0859	99.9141

COLUMN PROFILES

Reflux Ratio: 0.6034 Reboil Ratio: 0.5617 The Flows Option is Selected Flow Basis: Molar

Column Profiles Flows

	Temp (F)	Pres (psia)	Net Liq (lbmole/hr)	Net Vap (lbmole/hr)	Net Feed (lbmole/hr)	Net Draws (lbmole/hr)	Duty (Btu/hr)
Condenser	117.3	153.7	348.8	---	---	578.0	7.29e+006
1__Main Tower	117.6	153.7	345.5	926.8	---	---	---
2__Main Tower	118.8	153.8	335.4	923.5	---	---	---
3__Main Tower	122.5	153.9	306.3	913.4	---	---	---
4__Main Tower	133.5	154.0	248.8	884.4	---	---	---
5__Main Tower	156.7	154.2	195.7	826.9	---	---	---

6__Main Tower	179.6	154.3	170.6	773.8	---	---	---
7__Main Tower	192.3	154.4	155.6	748.6	---	---	---
8__Main Tower	202.1	154.5	1817	733.6	1668	---	---
9__Main Tower	204.6	154.6	1809	726.6	---	---	---
10__Main Tower	208.9	154.7	1797	719.0	---	---	---
11__Main Tower	216.3	154.8	1780	707.1	---	---	---
12__Main Tower	227.4	154.9	1761	690.3	---	---	---
13__Main Tower	241.9	155.0	1744	670.8	---	---	---
14__Main Tower	257.7	155.1	1736	654.2	---	---	---
15__Main Tower	273.1	155.2	1709	645.9	---	---	---
16__Main Tower	282.7	155.3	1707	619.2	---	---	---
17__Main Tower	289.4	155.4	1708	616.8	---	---	---
18__Main Tower	293.6	155.5	1710	618.2	---	---	---
19__Main Tower	296.1	155.6	1711	619.8	---	---	---
20__Main Tower	297.6	155.7	1711	620.9	---	---	---
21__Main Tower	298.5	155.8	1711	621.4	---	---	---
22__Main Tower	299.3	155.9	1709	621.1	---	---	---
23__Main Tower	300.5	156.0	1702	619.1	---	---	---
Reboiler	305.2	156.0	---	612.3	---	1090	8.45e+006

Column Profiles Energy

	Temperature (F)	Liq Enthalpy (Btu/lbmole)	Vap Enthalpy (Btu/lbmole)	Heat Loss (Btu/hr)
Condenser	117.3	-8.641e+004	-7.854e+004	---
1__Main Tower	117.6	-8.646e+004	-7.854e+004	---
2__Main Tower	118.8	-8.665e+004	-7.853e+004	---
3__Main Tower	122.5	-8.723e+004	-7.851e+004	---
4__Main Tower	133.5	-8.897e+004	-7.845e+004	---
5__Main Tower	156.7	-9.210e+004	-7.836e+004	---
6__Main Tower	179.6	-9.480e+004	-7.842e+004	---
7__Main Tower	192.3	-9.699e+004	-7.858e+004	---
8__Main Tower	202.1	-1.013e+005	-7.871e+004	---
9__Main Tower	204.6	-1.015e+005	-7.881e+004	---
10__Main Tower	208.9	-1.017e+005	-7.900e+004	---
11__Main Tower	216.3	-1.022e+005	-7.935e+004	---
12__Main Tower	227.4	-1.028e+005	-7.997e+004	---
13__Main Tower	241.9	-1.036e+005	-8.100e+004	---
14__Main Tower	257.7	-1.043e+005	-8.244e+004	---
15__Main Tower	273.1	-1.051e+005	-8.404e+004	---
16__Main Tower	282.7	-1.055e+005	-8.544e+004	---
17__Main Tower	289.4	-1.057e+005	-8.646e+004	---
18__Main Tower	293.6	-1.059e+005	-8.714e+004	---
19__Main Tower	296.1	-1.059e+005	-8.755e+004	---
20__Main Tower	297.6	-1.060e+005	-8.779e+004	---
21__Main Tower	298.5	-1.061e+005	-8.797e+004	---
22__Main Tower	299.3	-1.062e+005	-8.815e+004	---
23__Main Tower	300.5	-1.065e+005	-8.847e+004	---
Reboiler	305.2	-1.085e+005	-8.919e+004	---

FEEDS / PRODUCTS

Flow Basis: Molar

	Stream	Type	Duty (Btu/hr)	State	Flows (lbmole/hr)	Enthalpy (Btu/lbmole)	Temp (F)
Condenser	Q Condense 1	Energy	7.3e+006		---	---	---
	DME Outlet	Draw	---	Liquid	578	-8.6e+004	117.26
1__Main Tower							
2__Main Tower							
3__Main Tower							
4__Main Tower							
5__Main Tower							
6__Main Tower							
7__Main Tower							
8__Main Tower	Cooled Reactor Outlet Feed		---	Liquid	1.67e+003	-1.0e+005	201.96
9__Main Tower							
10__Main Tower							
11__Main Tower							
12__Main Tower							
13__Main Tower							
14__Main Tower							
15__Main Tower							
16__Main Tower							
17__Main Tower							
18__Main Tower							
19__Main Tower							
20__Main Tower							
21__Main Tower							
22__Main Tower							
23__Main Tower							
Reboiler	Q Reboil 1	Energy	8.4e+006		---	---	---
	Methanol and Water Outlet Draw		---	Liquid	1.09e+003	-1.1e+005	305.21

PERFORMANCE SUMMARY FOR INTERNAL OPTION: Internals-1@Main Tower@COL1

Number Of Stages	23
Total Height (ft)	46.00
Total Head Loss (in)	93.69
Total Pressure Drop (inH2O(60F))	61.56
Number Of Sections	1
Number Of Diameters	1
Pressure Drop Across Sump (psi)	---

Section	Start	End	Height (ft)	Diameter (ft)	Internals Type	Tray or Packing Type	Section Pressure Drop (inH2O(60F))	Approach To Flood (%)	Limiting Stage
CS-1	1__Main Tower	23__Main Tower	2.000	3.000	Trayed	Sieve	61.56	78.49	

SETUP

Sub-Flowsheet

Internal Feed Stream	External Feed Stream	Transfer Basis
Q Reboil 1	Q Reboil 1	None Req'd

Cooled Reactor Outlet	Cooled Reactor Outlet	P-H Flash
Internal Prod Stream	External Prod Stream	Transfer Basis
Q Condense 1	Q Condense 1	None Req'd
DME Outlet	DME Outlet	P-H Flash
Methanol and Water Outlet	Methanol and Water Outlet	P-H Flash

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
Q Reboil 1		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
Q Condense 1		
DME Outlet		

DYNAMICS

Vessel Dynamic Specifications

Vessel	Reboiler @COL1	Condenser @COL1
Diameter (ft)	3.914	3.914
Height.0 (ft)	5.871	5.871
Volume.0 (ft3)	70.63	70.63
Liquid Volume Percent (%)	50.00	50.00
Level Calculator	Horizontal cylinder	Horizontal cylinder
Fraction Calculator	Use levels and nozzles	Use levels and nozzles
Vessel Delta P (psi)	0.0000	0.0000
Fixed Vessel P Spec (psia)	156.0	153.7
Fixed P Spec Active	Not Active	Not Active

Other Equipment in Column Flowsheet

Holdup Details

	Pressure (psia)	Volume (ft3)	Bulk Liquid Volume (ft3)
Condenser	0.0000	0.0000	0.0000
1__Main Tower	0.0000	0.0000	0.0000
2__Main Tower	0.0000	0.0000	0.0000
3__Main Tower	0.0000	0.0000	0.0000
4__Main Tower	0.0000	0.0000	0.0000

5__Main Tower	0.0000	0.0000	0.0000
6__Main Tower	0.0000	0.0000	0.0000
7__Main Tower	0.0000	0.0000	0.0000
8__Main Tower	0.0000	0.0000	0.0000
9__Main Tower	0.0000	0.0000	0.0000
10__Main Tower	0.0000	0.0000	0.0000
11__Main Tower	0.0000	0.0000	0.0000
12__Main Tower	0.0000	0.0000	0.0000
13__Main Tower	0.0000	0.0000	0.0000
14__Main Tower	0.0000	0.0000	0.0000
15__Main Tower	0.0000	0.0000	0.0000
16__Main Tower	0.0000	0.0000	0.0000
17__Main Tower	0.0000	0.0000	0.0000
18__Main Tower	0.0000	0.0000	0.0000
19__Main Tower	0.0000	0.0000	0.0000
20__Main Tower	0.0000	0.0000	0.0000
21__Main Tower	0.0000	0.0000	0.0000
22__Main Tower	0.0000	0.0000	0.0000
23__Main Tower	0.0000	0.0000	0.0000
Reboiler	0.0000	0.0000	0.0000

T-101 (Distillation): Design, Profiles, Efficiencies, Solver, Internals, Rating, Worksheet, Performance, Flowsheet, Dynamics

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Distillation: T-101

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
Q Reboil 2	Reboiler	
Methanol and Water Outlet	19__Main Tower	T-100 Distillation

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
Q Condense 2	Condenser	
Methanol Out	Condenser	MIX-101 Mixer
Wastewater Out	Reboiler	V-100 Separator

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error
Reflux Ratio	1.500	1.019	-0.3209
Distillate Rate	749.6 lbmole/hr	507.7 lbmole/hr	-0.3226
Reflux Rate	---	517.2 lbmole/hr	---
Btms Prod Rate	582.2 lbmole/hr	582.2 lbmole/hr	-3.068e-007
Comp Fraction of Methanol in the top	0.9500	0.9389	-0.4118

Comp Fraction - 2	---	---	---

	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
Reflux Ratio	1.000e-002	1.000e-002	Off	On	Off
Distillate Rate	1.000e-002	2.205 lbmole/hr	Off	Off	Off
Reflux Rate	1.000e-002	2.205 lbmole/hr	Off	Off	Off
Btms Prod Rate	1.000e-002	2.205 lbmole/hr	On	On	On
Comp Fraction of Methanol in the top	1.000e-002	1.000e-003	Off	On	Off
Comp Fraction - 2	1.000e-002	1.000e-003	Off	On	Off

SPECS

Column Specification Parameters

Reflux Ratio

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Condenser Flow Basis: Molar Liquid Spec: ---

Distillate Rate

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Flow Basis: Liquid Spec:

Reflux Rate

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Flow Basis: Liquid Spec:

Btms Prod Rate

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stream: Wastewater Out @COL2 Flow Basis: Molar

Comp Fraction of Methanol in the top

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stage: Condenser Flow Basis: Mole Fraction Phase: Liquid

Components: Methanol

Comp Fraction - 2

Fix/Rang: Fixed Prim/Alter: Primary Lower Bnd: --- Upper Bnd: ---

Stream: Flow Basis:

Components:

Fix/Rang: Prim/Alter: Lower Bnd: Upper Bnd:

Draw: Flow Basis:

Components:

Fix/Rang:	Prim/Alter:	Lower Bnd:	Upper Bnd:
Stage:	Flow Basis:	Phase:	

Components:

SUBCOOLING

	Condenser
Degrees of Subcooling	---
Subcool to	---

User Variables

PROFILES

General Parameters

Sub-Flow Sheet: T-101 (COL2) Number of Stages: 32

Profile Estimates

	Temperature (F)	Net Liquid (lbmole/hr)	Net Vapour (lbmole/hr)
Condenser	269.2	517.2	1.012e-007
1__Main Tower	270.7	511.7	1025
2__Main Tower	272.0	506.3	1019
3__Main Tower	273.3	501.0	1014
4__Main Tower	274.7	495.9	1009
5__Main Tower	276.0	490.9	1004
6__Main Tower	277.4	486.2	998.7
7__Main Tower	278.8	481.6	993.9
8__Main Tower	280.2	477.2	989.3
9__Main Tower	281.5	473.1	984.9
10__Main Tower	282.9	469.3	980.8
11__Main Tower	284.1	465.9	977.1
12__Main Tower	285.3	462.9	973.7
13__Main Tower	286.4	460.4	970.7
14__Main Tower	287.4	458.2	968.1
15__Main Tower	288.2	456.3	965.9
16__Main Tower	288.9	454.9	964.1
17__Main Tower	289.5	453.7	962.6
18__Main Tower	290.1	452.1	961.4
19__Main Tower	291.3	1515	959.8
20__Main Tower	290.7	1513	932.8
21__Main Tower	290.8	1512	930.7
22__Main Tower	291.0	1512	930.2
23__Main Tower	291.4	1510	929.4
24__Main Tower	291.9	1508	928.0
25__Main Tower	292.9	1503	925.4
26__Main Tower	294.5	1496	920.9
27__Main Tower	297.4	1482	913.3
28__Main Tower	304.8	1457	899.7
29__Main Tower	314.5	1441	874.4
30__Main Tower	325.5	1431	858.5

31__Main Tower	335.2	1429	848.9
32__Main Tower	341.3	1429	846.7
Reboiler	344.8	582.2	847.0

EFFICIENCIES

Stage Efficiencies

Stages	Overall	H2O	diM-Ether	Methanol
Condenser	1.000	1.000	1.000	1.000
1__Main Tower	0.7200	0.7200	0.7200	0.7200
2__Main Tower	0.7200	0.7200	0.7200	0.7200
3__Main Tower	0.7200	0.7200	0.7200	0.7200
4__Main Tower	0.7200	0.7200	0.7200	0.7200
5__Main Tower	0.7200	0.7200	0.7200	0.7200
6__Main Tower	0.7200	0.7200	0.7200	0.7200
7__Main Tower	0.7200	0.7200	0.7200	0.7200
8__Main Tower	0.7200	0.7200	0.7200	0.7200
9__Main Tower	0.7200	0.7200	0.7200	0.7200
10__Main Tower	0.7200	0.7200	0.7200	0.7200
11__Main Tower	0.7200	0.7200	0.7200	0.7200
12__Main Tower	0.7200	0.7200	0.7200	0.7200
13__Main Tower	0.7200	0.7200	0.7200	0.7200
14__Main Tower	0.7200	0.7200	0.7200	0.7200
15__Main Tower	0.7200	0.7200	0.7200	0.7200
16__Main Tower	0.7200	0.7200	0.7200	0.7200
17__Main Tower	0.7200	0.7200	0.7200	0.7200
18__Main Tower	0.7200	0.7200	0.7200	0.7200
19__Main Tower	0.7200	0.7200	0.7200	0.7200
20__Main Tower	0.7200	0.7200	0.7200	0.7200
21__Main Tower	0.7200	0.7200	0.7200	0.7200
22__Main Tower	0.7200	0.7200	0.7200	0.7200
23__Main Tower	0.7200	0.7200	0.7200	0.7200
24__Main Tower	0.7200	0.7200	0.7200	0.7200
25__Main Tower	0.7200	0.7200	0.7200	0.7200
26__Main Tower	0.7200	0.7200	0.7200	0.7200
27__Main Tower	0.7200	0.7200	0.7200	0.7200
28__Main Tower	1.000	1.000	1.000	1.000
29__Main Tower	0.7200	0.7200	0.7200	0.7200
30__Main Tower	0.7200	0.7200	0.7200	0.7200
31__Main Tower	0.7200	0.7200	0.7200	0.7200
32__Main Tower	0.7200	0.7200	0.7200	0.7200
Reboiler	1.000	1.000	1.000	1.000

SOLVER

Column Solving Algorithm: Modified HYSIM Inside-Out

Solving Options	Acceleration Parameters
Maximum Iterations: 10000	Accelerate K Value & H Model Parameters: Off
Equilibrium Error Tolerance: 1.000e-05	
Heat/Spec Error Tolerance: 5.000e-004	

Save Solutions as Initial Estimate: On

Super Critical Handling Model: Simple K

Trace Level: Low

Init from Ideal K's: Off Damping Parameters

Initial Estimate Generator Parameters Azeotrope Check: Off

Iterative IEG (Good for Chemicals): Off Fixed Damping Factor: 1

ACTIVE INTERNAL OPTION: Internals-1@Main Tower@COL2

Name	Start Stage	End Stage	Mode	Internals	Tray / Packing Type	Number of Passes	Packing Vendor	Packing Material	Packing Dimension	Tray Spacing / Section Packed Height (ft)	Diameter (ft)
CS-1	1__Main Tower	32__Main Tower	Interactive Sizing	Trayed	Sieve	1	---	---	---	2.000	3.000

SETUP

Section Name CS-1

Section Start 1__Main Tower

Section End 32__Main Tower

Internals Trayed

Internals Type Sieve

Diameter (ft) 3.000

Tray Spacing / Section Packed Height (ft) 2.000

Number Of Passes 1

Maximum Acceptable Pressure Drop (psi) 0.3626

Maximum Percent Downcomer Backup 100.00 %

Maximum Percent Jet Flood 100.00 %

Percent Jet Flood For Design 80.00 %

Maximum Percent Liquid Entrainment 10.00 %

Minimum Weir Loading (USGPM/ft) 6.000

Maximum Weir Loading (USGPM/ft) 157.5

Minimum Downcomer Area / Total Tray Area 0.1000

Override Downcomer Froth Density No

Froth Density ---

Weep Method Hsieh

Default Jet Flood Calculation Method GLITSCH6

Maximum Downcomer Loading Method Glitsch

% Approach to Maximum Capacity ---

Design Capacity Factor ---

Capacity Factor at Flooding ---

System Foaming Factor 1.000

Aeration Factor Multipler 1.000

Minimum Liquid Flow Rate ---

Pressure Drop at Flood per Unit Packed Height ---

Allowable Pressure Drop per Unit Packed Height ---

Minimum Pressure Drop per Unit Packed Height ---

Number of Curves ---

Warning Status (% to Limit) 10.00 %

Pressure Drop Calculation Method ---

Mode Interactive Sizing

Status Needs Calculating

GEOMETRY DETAILS

Common Geometry CS-1

Section Start 1__Main Tower

Section End 32__Main Tower

Internals Sieve

Section Diameter (ft) 3.000

Foaming Factor 1.000

Over-Design Factor 1.000

Common Tray Geometry CS-1

Number of Passes 1

Tray Spacing (ft) 2.000

Picket Fence Weirs No

Swept Back Weirs No

Active Area Under Downcomer No

Deck Thickness 10 Gauge

Deck Thickness Value (in) 0.1340

Balance Downcomers Based On Maximum Downcomer Loading

Weir Modifications None

Net Area (ft2) 6.362

Cross-Sectional Area (ft2) 7.069

Active Area (ft2) 5.655

Downcomer Geometry CS-1

Side Weir Height (in) 2.000

Weir Length (ft) ---

Downcomer Clearance (in) 1.500

Downcomer Width - Top (in) 5.633

Downcomer Width - Bottom (in) 5.633

Downcomer Loading Top (USGPM/ft2) 66.00

Weir Loading (USGPM/ft) 44.69

Downcomer Area - Top (ft2) 0.7069

Downcomer Area - Bottom (ft2) 0.7069

Picketing Fraction ---

Center Weir Height (in) ---

Weir Length (ft) ---

Downcomer Clearance (in) ---

Downcomer Width - Top (in) ---

Downcomer Width - Bottom (in) ---

Downcomer Loading Top (USGPM/ft2) ---

Weir Loading (USGPM/ft) ---

Downcomer Area - Top (ft2) ---

Downcomer Area - Bottom (ft2) ---

Picketing Fraction ---

Off Center Weir Height (in) ---

Inside Weir Length (ft) ---

Outside Weir Length (ft) ---

Downcomer Clearance (in) ---

Downcomer Width - Top (in) ---

Downcomer Width - Bottom (in) ---

Downcomer Loading Top (USGPM/ft2) ---

Maximum Outside Weir Loading (USGPM/ft) ---

Maximum Inside Weir Loading (USGPM/ft) ---

Downcomer Area - Top (ft2) ---

Downcomer Area - Bottom (ft2) ---

Inside Picketing Fraction ---

Outside Picketing Fraction ---

Off-Center Downcomer Location (ft) ---

Swept Back Weir Geometry CS-1

Compatibility KG Tower

A ---

B/Parallel Chord Segment ---

S/Swept-Back Weir ---

Swept-Back Weir Chord ---

Angled Chord Segment ---

Tray With Maximum Weir Loading 19

Maximum Weir Loading (USGPM/ft) 44.69

Maximum Allowable Weir Loading in Section (USGPM/ft) 157.5

Actual Side Weir Length (ft) 2.180

Effective Side Weir Length (ft) 2.180

Lost Area (%) 0.00

Sieve Geometry CS-1

Hole Diameter (in) 0.5000

Number of Holes 414

Hole Area to Active Area 0.1000

Bubble Cap Geometry CS-1

Cap Diameter ---

Skirt Height ---

Number of Caps ---

Number of Caps Per Active Area ---

Valve Geometry CS-1

Tray Type ---

Valve Type ---

Valve Material ---

Leg Length ---

Valve Thickness ---

Number of Valves ---

Number of Valves per Active Area ---

Packing Geometry CS-1

HETP (ft) ---

Section Packed Height (ft) ---

Packing Type ---

Packing Vendor ---

Packing Material ---

Packing Dimension ---

Packing Factor (ft2/ft3) ---

Packing Surface Area (ft2/ft3) ---

1st Stichlmair Constant ---

2nd Stichlmair Constant ---

3rd Stichlmair Constant ---

Void Fraction ---

RESULTS SUMMARY

Section Name CS-1

Section Start 1__Main Tower

Section End 32__Main Tower

Internals Trayed

Diameter (ft) 3.000

Number of Passes 1

Tray Spacing / Section Packed Height (ft) 2.000

Total Height (ft) 64.00

Total Pressure Drop (psi) 99.06

Total Pressure Drop (Head Loss) (ft) 133.8

Trays With Weeping None

Maximum Percent Jet Flood (%) 79.66

Tray With Maximum Jet Flood 1__Main Tower

Maximum Percent Downcomer Backup (%) 36.08

Tray With Maximum Downcomer Backup 1__Main Tower

Maximum Downcomer Loading (USGPM/ft2) 137.8

Tray With Maximum Downcomer Loading 19__Main Tower

Maximum Downcomer Loading Location Side

Maximum Weir Loading (USGPM/ft) 44.69

Tray With Maximum Weir Loading 19__Main Tower

Maximum Weir Loading Location Side

Maximum Aerated Height Over Weir (in) 5.329

Tray With Maximum Aerated Height Over Weir 19__Main Tower

Maximum % Approach To System Limit (%) 56.46

Tray With Maximum % Approach To System Limit 1__Main Tower

Maximum Cs Based On Bubbling Area (%) 0.3431

Tray With Maximum Cs Based On Bubbling Area 1__Main Tower

Maximum % Capacity (Constant L/V) 79.66

Maximum Capacity Factor ---

Section Pressure Drop (psi) 99.06

Average Pressure Drop Per Height (inH2O/ft) ---

Average Pressure Drop Per Height (Frictional) (inH2O/ft) ---

Maximum Stage Liquid Holdup (ft3) ---

Maximum Liquid Superficial Velocity (ft/s) ---

Surface Area (ft2/ft3) ---

Void Fraction ---

1st Stichlmair Constant ---

2nd Stichlmair Constant ---

3rd Stichlmair Constant ---

STAGE BY STAGE RESULTS: CS-1

State Conditions

Stages	Liquid Temperature (F)	Vapor Temperature (F)	Liquid Mass Flow (lb/hr)	Vapor Mass Flow (lb/hr)	Liquid Volume Flow (USGPS)	Vapor Volume Flow (USGPS)
1__Main Tower	270.7	272.0	1.568e+004	3.152e+004	0.7776	133.2
2__Main Tower	272.0	273.3	1.525e+004	3.109e+004	0.7507	132.6
3__Main Tower	273.3	274.7	1.483e+004	3.067e+004	0.7247	132.1
4__Main Tower	274.7	276.0	1.442e+004	3.026e+004	0.6995	131.5
5__Main Tower	276.0	277.4	1.403e+004	2.987e+004	0.6751	131.0
6__Main Tower	277.4	278.8	1.364e+004	2.948e+004	0.6515	130.5
7__Main Tower	278.8	280.2	1.328e+004	2.912e+004	0.6290	130.0
8__Main Tower	280.2	281.5	1.293e+004	2.877e+004	0.6076	129.5
9__Main Tower	281.5	282.9	1.260e+004	2.844e+004	0.5876	129.1
10__Main Tower	282.9	284.1	1.230e+004	2.814e+004	0.5692	128.7
11__Main Tower	284.1	285.3	1.203e+004	2.787e+004	0.5526	128.3
12__Main Tower	285.3	286.4	1.179e+004	2.763e+004	0.5380	128.0
13__Main Tower	286.4	287.4	1.159e+004	2.743e+004	0.5253	127.7
14__Main Tower	287.4	288.2	1.141e+004	2.725e+004	0.5146	127.4
15__Main Tower	288.2	288.9	1.126e+004	2.710e+004	0.5057	127.2
16__Main Tower	288.9	289.5	1.114e+004	2.698e+004	0.4984	127.0
17__Main Tower	289.5	290.1	1.105e+004	2.689e+004	0.4926	126.8
18__Main Tower	290.1	291.3	1.096e+004	2.680e+004	0.4873	126.7
19__Main Tower	291.3	290.7	3.656e+004	2.603e+004	1.623	123.0
20__Main Tower	290.7	290.8	3.648e+004	2.595e+004	1.619	122.5
21__Main Tower	290.8	291.0	3.643e+004	2.590e+004	1.616	122.4
22__Main Tower	291.0	291.4	3.636e+004	2.583e+004	1.611	122.2
23__Main Tower	291.4	291.9	3.622e+004	2.569e+004	1.603	122.0
24__Main Tower	291.9	292.9	3.598e+004	2.545e+004	1.588	121.7
25__Main Tower	292.9	294.5	3.557e+004	2.504e+004	1.563	121.3
26__Main Tower	294.5	297.4	3.488e+004	2.435e+004	1.521	120.6
27__Main Tower	297.4	304.8	3.373e+004	2.320e+004	1.452	119.9
28__Main Tower	304.8	314.5	3.116e+004	2.063e+004	1.296	117.9
29__Main Tower	314.5	325.5	2.916e+004	1.863e+004	1.176	117.3
30__Main Tower	325.5	335.2	2.752e+004	1.699e+004	1.079	117.3
31__Main Tower	335.2	341.3	2.655e+004	1.602e+004	1.021	117.8
32__Main Tower	341.3	344.8	2.609e+004	1.556e+004	0.9936	118.3

Physical Conditions

Stages	Liquid Molecular Weight	Vapor Molecular Weight	Liquid Mass Density (lb/ft3)	Vapor Mass Density (lb/ft3)	Liquid Viscosity (cP)	Vapor Viscosity (cP)	Surface Tension (dyne/cm)
1__Main Tower	30.65	30.92	41.92	0.4917	0.1626	8.630e-003	17.07
2__Main Tower	30.12	30.66	42.21	0.4871	0.1627	8.683e-003	18.43
3__Main Tower	29.60	30.40	42.52	0.4826	0.1628	8.736e-003	19.78
4__Main Tower	29.08	30.15	42.84	0.4782	0.1628	8.788e-003	21.10
5__Main Tower	28.57	29.91	43.17	0.4738	0.1628	8.840e-003	22.40
6__Main Tower	28.06	29.67	43.51	0.4696	0.1627	8.892e-003	23.68
7__Main Tower	27.57	29.43	43.86	0.4655	0.1624	8.943e-003	24.91

8__Main Tower	27.10	29.21	44.22	0.4615	0.1622	8.993e-003	26.10
9__Main Tower	26.64	29.00	44.57	0.4578	0.1618	9.041e-003	27.23
10__Main Tower	26.22	28.81	44.92	0.4544	0.1613	9.085e-003	28.28
11__Main Tower	25.83	28.63	45.25	0.4513	0.1608	9.127e-003	29.24
12__Main Tower	25.48	28.47	45.55	0.4486	0.1603	9.165e-003	30.09
13__Main Tower	25.17	28.33	45.83	0.4462	0.1597	9.198e-003	30.83
14__Main Tower	24.91	28.21	46.07	0.4443	0.1592	9.227e-003	31.46
15__Main Tower	24.68	28.11	46.28	0.4427	0.1586	9.251e-003	31.99
16__Main Tower	24.50	28.03	46.46	0.4415	0.1582	9.271e-003	32.42
17__Main Tower	24.35	27.97	46.60	0.4406	0.1577	9.288e-003	32.76
18__Main Tower	24.23	27.92	46.72	0.4395	0.1573	9.313e-003	33.04
19__Main Tower	24.13	27.90	46.79	0.4396	0.1563	9.305e-003	33.20
20__Main Tower	24.11	27.88	46.84	0.4400	0.1569	9.310e-003	33.31
21__Main Tower	24.09	27.85	46.86	0.4398	0.1568	9.318e-003	33.35
22__Main Tower	24.05	27.79	46.89	0.4391	0.1566	9.332e-003	33.44
23__Main Tower	23.99	27.69	46.96	0.4375	0.1563	9.355e-003	33.59
24__Main Tower	23.87	27.51	47.08	0.4345	0.1559	9.394e-003	33.85
25__Main Tower	23.67	27.19	47.28	0.4290	0.1551	9.463e-003	34.31
26__Main Tower	23.32	26.66	47.65	0.4194	0.1537	9.581e-003	35.08
27__Main Tower	22.76	25.79	48.27	0.4021	0.1510	9.817e-003	36.32
28__Main Tower	21.39	23.59	49.96	0.3637	0.1425	1.028e-002	39.24
29__Main Tower	20.24	21.70	51.52	0.3301	0.1328	1.071e-002	41.29
30__Main Tower	19.23	20.02	53.02	0.3010	9.700e-002	1.106e-002	42.73
31__Main Tower	18.58	18.92	54.04	0.2827	9.040e-002	1.125e-002	43.30
32__Main Tower	18.25	18.37	54.56	0.2732	9.030e-002	1.134e-002	43.43

Hydraulic Results

Stages	Percent Jet Flood	Dry Pressure Drop	Total Pressure Drop	Dry Pressure Drop (Head Loss)	Total Pressure Drop (Head Loss)
	(%)	(inH2O(60F))	(inH2O(60F))	(in)	(in)
1__Main Tower	79.66	2.689	3.458	4.002	5.145
2__Main Tower	78.56	2.640	3.417	3.901	5.049
3__Main Tower	77.49	2.593	3.379	3.804	4.956
4__Main Tower	76.43	2.548	3.343	3.710	4.866
5__Main Tower	75.40	2.505	3.308	3.619	4.779
6__Main Tower	74.39	2.463	3.275	3.531	4.694
7__Main Tower	73.41	2.424	3.244	3.446	4.613
8__Main Tower	72.47	2.386	3.216	3.366	4.536
9__Main Tower	71.58	2.352	3.190	3.290	4.463
10__Main Tower	70.75	2.320	3.166	3.221	4.396
11__Main Tower	70.00	2.291	3.145	3.157	4.335
12__Main Tower	69.32	2.265	3.127	3.101	4.281
13__Main Tower	68.72	2.243	3.111	3.052	4.233
14__Main Tower	68.21	2.224	3.098	3.010	4.193
15__Main Tower	67.78	2.208	3.087	2.975	4.159
16__Main Tower	67.42	2.195	3.077	2.946	4.131
17__Main Tower	67.12	2.184	3.070	2.922	4.108
18__Main Tower	66.87	2.174	3.063	2.902	4.088
19__Main Tower	70.04	2.051	3.187	2.733	4.248
20__Main Tower	69.78	2.037	3.176	2.712	4.229

21__Main Tower	69.66	2.031	3.171	2.703	4.221
22__Main Tower	69.48	2.022	3.164	2.689	4.208
23__Main Tower	69.17	2.008	3.152	2.667	4.187
24__Main Tower	68.66	1.984	3.133	2.629	4.150
25__Main Tower	67.79	1.945	3.100	2.566	4.089
26__Main Tower	66.33	1.881	3.048	2.462	3.990
27__Main Tower	64.00	1.781	2.967	2.301	3.833
28__Main Tower	58.55	1.558	2.796	1.944	3.490
29__Main Tower	54.42	1.399	2.682	1.694	3.247
30__Main Tower	51.04	1.276	2.602	1.501	3.060
31__Main Tower	49.08	1.209	2.561	1.395	2.955
32__Main Tower	48.18	1.179	2.544	1.348	2.908

Stages	Downcomer Backup (Aerated)	Downcomer Backup (Unaerated)	Percent Downcomer Backup (Aerated)	Percent Downcomer Backup (Unaerated)
	(ft)	(ft)	(%)	(%)
1__Main Tower	0.7818	0.4719	36.08	21.78
2__Main Tower	0.7683	0.4640	35.46	21.42
3__Main Tower	0.7553	0.4564	34.86	21.06
4__Main Tower	0.7428	0.4490	34.28	20.72
5__Main Tower	0.7307	0.4419	33.72	20.40
6__Main Tower	0.7190	0.4350	33.18	20.08
7__Main Tower	0.7079	0.4285	32.67	19.78
8__Main Tower	0.6973	0.4223	32.18	19.49
9__Main Tower	0.6875	0.4164	31.73	19.22
10__Main Tower	0.6784	0.4111	31.31	18.97
11__Main Tower	0.6703	0.4062	30.93	18.75
12__Main Tower	0.6630	0.4019	30.60	18.55
13__Main Tower	0.6567	0.3982	30.31	18.38
14__Main Tower	0.6514	0.3950	30.06	18.23
15__Main Tower	0.6469	0.3924	29.86	18.11
16__Main Tower	0.6432	0.3902	29.68	18.01
17__Main Tower	0.6401	0.3884	29.55	17.92
18__Main Tower	0.6376	0.3868	29.43	17.85
19__Main Tower	0.7525	0.4566	34.73	21.07
20__Main Tower	0.7499	0.4550	34.61	21.00
21__Main Tower	0.7487	0.4543	34.55	20.97
22__Main Tower	0.7468	0.4532	34.47	20.92
23__Main Tower	0.7437	0.4513	34.33	20.83
24__Main Tower	0.7385	0.4481	34.08	20.68
25__Main Tower	0.7296	0.4428	33.67	20.44
26__Main Tower	0.7152	0.4341	33.01	20.04
27__Main Tower	0.6927	0.4206	31.97	19.41
28__Main Tower	0.6445	0.3916	29.75	18.07
29__Main Tower	0.6115	0.3716	28.22	17.15
30__Main Tower	0.5867	0.3567	27.08	16.46
31__Main Tower	0.5731	0.3485	26.45	16.08
32__Main Tower	0.5669	0.3447	26.17	15.91

Stages Mass Rate / Column Area Volume Rate / Column Area Fs (Net Area) Fs (Bubble Area) Cs (Net Area)

	(lb/s-ft2)	(USGPM/ft2)	(ft/(s/sqrt(lb/ft3)))	(ft/(s/sqrt(lb/ft3)))	(ft/s)
1__Main Tower	0.6164	6.600	1.963	2.208	0.3050
2__Main Tower	0.5993	6.372	1.945	2.188	0.3011
3__Main Tower	0.5828	6.151	1.928	2.169	0.2973
4__Main Tower	0.5667	5.938	1.911	2.150	0.2936
5__Main Tower	0.5512	5.730	1.894	2.131	0.2899
6__Main Tower	0.5362	5.530	1.879	2.113	0.2863
7__Main Tower	0.5218	5.339	1.863	2.096	0.2829
8__Main Tower	0.5081	5.157	1.849	2.080	0.2795
9__Main Tower	0.4953	4.988	1.836	2.065	0.2764
10__Main Tower	0.4835	4.832	1.823	2.051	0.2734
11__Main Tower	0.4729	4.691	1.812	2.038	0.2707
12__Main Tower	0.4635	4.567	1.802	2.027	0.2682
13__Main Tower	0.4553	4.459	1.793	2.017	0.2661
14__Main Tower	0.4484	4.368	1.785	2.008	0.2643
15__Main Tower	0.4427	4.293	1.779	2.001	0.2627
16__Main Tower	0.4379	4.231	1.773	1.995	0.2614
17__Main Tower	0.4341	4.181	1.769	1.990	0.2603
18__Main Tower	0.4306	4.136	1.765	1.986	0.2594
19__Main Tower	1.437	13.78	1.714	1.928	0.2518
20__Main Tower	1.434	13.74	1.708	1.922	0.2508
21__Main Tower	1.432	13.71	1.706	1.919	0.2503
22__Main Tower	1.429	13.68	1.702	1.915	0.2497
23__Main Tower	1.423	13.61	1.696	1.908	0.2487
24__Main Tower	1.414	13.48	1.686	1.897	0.2469
25__Main Tower	1.398	13.27	1.669	1.878	0.2439
26__Main Tower	1.371	12.91	1.642	1.847	0.2389
27__Main Tower	1.325	12.33	1.598	1.797	0.2309
28__Main Tower	1.225	11.00	1.494	1.681	0.2121
29__Main Tower	1.146	9.984	1.416	1.593	0.1979
30__Main Tower	1.082	9.156	1.352	1.521	0.1863
31__Main Tower	1.043	8.667	1.316	1.480	0.1795
32__Main Tower	1.025	8.434	1.300	1.462	0.1764

Stages	Cs (Bubble Area)	Approach to System Limit	Height Over Weir (Aerated)	Height Over Weir (Unaerated)
	(ft/s)	(%)	(ft)	(ft)
1__Main Tower	0.3431	56.46	0.2289	4.101e-002
2__Main Tower	0.3388	54.74	0.2203	3.964e-002
3__Main Tower	0.3345	53.16	0.2123	3.835e-002
4__Main Tower	0.3303	51.71	0.2048	3.714e-002
5__Main Tower	0.3262	50.37	0.1978	3.600e-002
6__Main Tower	0.3221	49.12	0.1911	3.492e-002
7__Main Tower	0.3182	47.97	0.1848	3.390e-002
8__Main Tower	0.3145	46.92	0.1789	3.294e-002
9__Main Tower	0.3109	45.96	0.1735	3.206e-002
10__Main Tower	0.3076	45.10	0.1686	3.125e-002
11__Main Tower	0.3045	44.34	0.1641	3.053e-002
12__Main Tower	0.3018	43.68	0.1603	2.989e-002
13__Main Tower	0.2994	43.12	0.1569	2.934e-002

14__Main Tower	0.2973	42.64	0.1541	2.888e-002
15__Main Tower	0.2955	42.25	0.1518	2.850e-002
16__Main Tower	0.2941	41.93	0.1499	2.818e-002
17__Main Tower	0.2929	41.67	0.1484	2.793e-002
18__Main Tower	0.2919	41.46	0.1470	2.770e-002
19__Main Tower	0.2832	40.42	0.4441	8.500e-002
20__Main Tower	0.2821	40.24	0.4426	8.488e-002
21__Main Tower	0.2816	40.15	0.4419	8.480e-002
22__Main Tower	0.2809	40.03	0.4407	8.468e-002
23__Main Tower	0.2798	39.83	0.4386	8.444e-002
24__Main Tower	0.2778	39.47	0.4350	8.402e-002
25__Main Tower	0.2744	38.88	0.4289	8.329e-002
26__Main Tower	0.2688	37.90	0.4185	8.204e-002
27__Main Tower	0.2598	36.36	0.4015	7.996e-002
28__Main Tower	0.2386	32.90	0.3622	7.514e-002
29__Main Tower	0.2226	30.42	0.3113	6.685e-002
30__Main Tower	0.2095	28.49	0.2592	5.738e-002
31__Main Tower	0.2019	27.43	0.2389	5.390e-002
32__Main Tower	0.1985	26.97	0.2303	5.240e-002

Side Downcomer Results

Stages	Volume (ft3)	Residence Time (seconds)	Velocity From Top (ft/s)	Velocity from Bottom (ft/s)	Exit Velocity (ft/s)
1__Main Tower	0.3336	3.209	0.1471	0.1471	0.3815
2__Main Tower	0.3280	3.268	0.1420	0.1420	0.3683
3__Main Tower	0.3226	3.330	0.1371	0.1371	0.3555
4__Main Tower	0.3174	3.394	0.1323	0.1323	0.3432
5__Main Tower	0.3124	3.461	0.1277	0.1277	0.3312
6__Main Tower	0.3075	3.531	0.1232	0.1232	0.3196
7__Main Tower	0.3029	3.602	0.1189	0.1189	0.3086
8__Main Tower	0.2985	3.675	0.1149	0.1149	0.2981
9__Main Tower	0.2944	3.748	0.1111	0.1111	0.2883
10__Main Tower	0.2906	3.819	0.1076	0.1076	0.2793
11__Main Tower	0.2872	3.887	0.1045	0.1045	0.2711
12__Main Tower	0.2841	3.951	0.1017	0.1017	0.2639
13__Main Tower	0.2815	4.008	9.935e-002	9.935e-002	0.2577
14__Main Tower	0.2792	4.059	9.732e-002	9.732e-002	0.2525
15__Main Tower	0.2773	4.103	9.564e-002	9.564e-002	0.2481
16__Main Tower	0.2758	4.139	9.426e-002	9.426e-002	0.2445
17__Main Tower	0.2745	4.169	9.316e-002	9.316e-002	0.2417
18__Main Tower	0.2734	4.198	9.216e-002	9.216e-002	0.2391
19__Main Tower	0.3227	1.487	0.3070	0.3070	0.7965
20__Main Tower	0.3216	1.486	0.3061	0.3061	0.7941
21__Main Tower	0.3211	1.487	0.3056	0.3056	0.7927
22__Main Tower	0.3203	1.487	0.3047	0.3047	0.7904
23__Main Tower	0.3190	1.489	0.3031	0.3031	0.7864
24__Main Tower	0.3168	1.492	0.3004	0.3004	0.7793
25__Main Tower	0.3130	1.498	0.2956	0.2956	0.7670
26__Main Tower	0.3069	1.509	0.2877	0.2877	0.7463

27__Main Tower	0.2973	1.531	0.2746	0.2746	0.7124
28__Main Tower	0.2768	1.597	0.2451	0.2451	0.6359
29__Main Tower	0.2627	1.671	0.2224	0.2224	0.5770
30__Main Tower	0.2521	1.749	0.2040	0.2040	0.5292
31__Main Tower	0.2463	1.805	0.1931	0.1931	0.5010
32__Main Tower	0.2437	1.835	0.1879	0.1879	0.4875

RATING

Tray Sections

Tray Section	Main Tower @COL2
Tray Diameter (ft)	4.921
Weir Height (ft)	0.1640
Weir Length (ft)	3.937
Tray Space (ft)	1.804
Tray Volume (ft3)	34.32
Disable Heat Loss Calculations	No
Heat Model	None
Rating Calculations	No
Tray Hold Up (ft3)	3.120

Vessels

Vessel	Reboiler @COL2	Condenser @COL2
Diameter (ft)	3.914	3.914
Length (ft)	5.871	5.871
Volume (ft3)	70.63	70.63
Orientation	Horizontal	Horizontal
Vessel has a Boot	No	No
Boot Diameter (ft)	---	---
Boot Length (ft)	---	---
Hold Up (ft3)	35.31	35.31

Other Equipment In Column Flowsheet

Pressure Profile

	Pressure (psia)	Pressure Drop (psi)
Condenser	124.7 psia	0.0000 psi
1__Main Tower	124.7 psia	0.1183 psi
2__Main Tower	124.9 psia	0.1183 psi
3__Main Tower	125.0 psia	0.1183 psi
4__Main Tower	125.1 psia	0.1183 psi
5__Main Tower	125.2 psia	0.1183 psi
6__Main Tower	125.3 psia	0.1183 psi
7__Main Tower	125.4 psia	0.1183 psi
8__Main Tower	125.6 psia	0.1183 psi
9__Main Tower	125.7 psia	0.1183 psi
10__Main Tower	125.8 psia	0.1183 psi

11__Main Tower	125.9 psia	0.1183 psi
12__Main Tower	126.0 psia	0.1183 psi
13__Main Tower	126.2 psia	0.1183 psi
14__Main Tower	126.3 psia	0.1183 psi
15__Main Tower	126.4 psia	0.1183 psi
16__Main Tower	126.5 psia	0.1183 psi
17__Main Tower	126.6 psia	0.1183 psi
18__Main Tower	126.7 psia	0.1183 psi
19__Main Tower	126.9 psia	0.1183 psi
20__Main Tower	127.0 psia	0.1183 psi
21__Main Tower	127.1 psia	0.1183 psi
22__Main Tower	127.2 psia	0.1183 psi
23__Main Tower	127.3 psia	0.1183 psi
24__Main Tower	127.5 psia	0.1183 psi
25__Main Tower	127.6 psia	0.1183 psi
26__Main Tower	127.7 psia	0.1183 psi
27__Main Tower	127.8 psia	0.1183 psi
28__Main Tower	127.9 psia	0.1183 psi
29__Main Tower	128.0 psia	0.1183 psi
30__Main Tower	128.2 psia	0.1183 psi
31__Main Tower	128.3 psia	0.1183 psi
32__Main Tower	128.4 psia	---
Reboiler	128.4 psia	0.0000 psi

Pressure Solving Options

Pressure Tolerance 1.000e-004 Pressure Drop Tolerance 1.000e-004

Damping Factor 1.000 Max Press Iterations 100

CONDITIONS

Name	Methanol and Water Outlet	Wastewater Out	Methanol Out	Q Reboil 2	Q Condense 2
Vapour	0.0000	0.0000	0.0000	---	---
Temperature (F)	305.2081	344.8441	269.1701	---	---
Pressure (psia)	156.0000	128.4000	124.7324	---	---
Molar Flow (lbmole/hr)	1089.9542	582.2347	507.7195	---	---
Mass Flow (lb/hr)	26369.6753	10529.8515	15839.8238	---	---
Std Ideal Liq Vol Flow (barrel/day)	2077.6358	724.0906	1353.5452	---	---
Molar Enthalpy (Btu/lbmole)	-1.085e+005	-1.175e+005	-9.858e+004	---	---
Molar Entropy (Btu/lbmole-F)	21.07	9.146	30.69	---	---
Heat Flow (Btu/hr)	-1.1829e+08	-6.8400e+07	-5.0049e+07	1.3358e+07	1.3520e+07

PROPERTIES

Name	Methanol and Water Outlet	Wastewater Out	Methanol Out
Molecular Weight	24.19	18.09	31.20
Molar Density (lbmole/ft3)	1.907	3.030	1.334
Mass Density (lb/ft3)	46.14	54.80	41.63
Act. Volume Flow (barrel/day)	2443	821.4	1626
Mass Enthalpy (Btu/lb)	-4486	-6496	-3160
Mass Entropy (Btu/lb-F)	0.8709	0.5057	0.9836

Heat Capacity (Btu/lbmole-F)	25.61	19.01	31.67
Mass Heat Capacity (Btu/lb-F)	1.059	1.051	1.015
LHV Molar Basis (Std) (Btu/lbmole)	1.208e+005	1372	2.578e+005
HHV Molar Basis (Std) (Btu/lbmole)	1.462e+005	1.909e+004	2.920e+005
HHV Mass Basis (Std) (Btu/lb)	6044	1056	9361
CO2 Loading	---	---	---
CO2 Apparent Mole Conc. (lbmole/ft3)	---	---	---
CO2 Apparent Wt. Conc. (lbmol/lb)	---	---	---
LHV Mass Basis (Std) (Btu/lb)	4995	75.85	8265
Phase Fraction [Vol. Basis]	0.0000	2.729e-006	0.0000
Phase Fraction [Mass Basis]	0.0000	2.705e-006	0.0000
Phase Fraction [Act. Vol. Basis]	0.0000	5.426e-004	0.0000
Mass Exergy (Btu/lb)	-87.38	49.92	-160.4
Partial Pressure of CO2 (psia)	0.0000	0.0000	0.0000
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000
Act. Gas Flow (ACFM)	---	1.738e-003	---
Avg. Liq. Density (lbmole/ft3)	2.243	3.437	1.603
Specific Heat (Btu/lbmole-F)	25.61	19.01	31.67
Std. Gas Flow (MMSCFD)	9.908	5.293	4.615
Std. Ideal Liq. Mass Density (lb/ft3)	54.25	62.16	50.02
Act. Liq. Flow (USGPS)	1.188	0.3991	0.7906
Z Factor	9.966e-003	---	1.195e-002
Watson K	10.63	10.63	10.63
User Property	---	---	---
Partial Pressure of H2S (psia)	0.0000	0.0000	0.0000
Cp/(Cp - R)	1.084	1.117	1.067
Cp/Cv	1.233	1.115	1.381
Heat of Vap. (Btu/lbmole)	1.474e+004	1.575e+004	1.320e+004
Kinematic Viscosity (cSt)	0.1957	---	0.2437
Liq. Mass Density (Std. Cond) (lb/ft3)	54.76	63.21	50.08
Liq. Vol. Flow (Std. Cond) (barrel/day)	2058	712.1	1352
Liquid Fraction	1.000	1.000	1.000
Molar Volume (ft3/lbmole)	0.5244	0.3301	0.7494
Mass Heat of Vap. (Btu/lb)	609.1	870.9	423.0
Phase Fraction [Molar Basis]	0.0000	0.0000	0.0000
Surface Tension (dyne/cm)	31.66	43.44	15.68
Thermal Conductivity (Btu/hr-ft-F)	0.2087	0.3918	8.659e-002
Viscosity (cP)	0.1447	9.011e-002	0.1625
Cv (Semi-Ideal) (Btu/lbmole-F)	23.62	17.02	29.68
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.9765	0.9413	0.9513
Cv (Btu/lbmole-F)	20.78	17.04	22.93
Mass Cv (Btu/lb-F)	0.8588	0.9424	0.7349
Cv (Ent. Method) (Btu/lbmole-F)	---	---	---
Mass Cv (Ent. Method) (Btu/lb-F)	---	---	---
Cp/Cv (Ent. Method)	---	---	---
Reid VP at 37.8 C (psia)	4.697	4.641	4.697
True VP at 37.8 C (psia)	3.173	1.003	4.525
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	2058	712.1	1352

Viscosity Index	---	---	-34.84

SUMMARY

Flow Basis: Molar The composition option is selected

Feed Composition

	Methanol and Water Outlet
Flow Rate (lbmole/hr)	1.089954e+03

H2O	0.5598
diM-Ether	0.0002
Methanol	0.4400

Flow Basis: Molar The composition option is selected

Feed Flows

	Methanol and Water Outlet
Flow Rate (lbmole/hr)	1.089954e+03

H2O (lbmole/hr)	610.1124
diM-Ether (lbmole/hr)	0.2421
Methanol (lbmole/hr)	479.5998

Products

Flow Basis: Molar The composition option is selected

Product Compositions

	Methanol Out	Wastewater Out
Flow Rate (lbmole/hr)	507.7195	582.2347
	---	---
H2O	0.0606	0.9950
diM-Ether	0.0005	0.0000
Methanol	0.9389	0.0050

Flow Basis: Molar The composition option is selected

Product Flows

	Methanol Out	Wastewater Out
Flow Rate (lbmole/hr)	507.7195	582.2347
	---	---
H2O (lbmole/hr)	30.7889	579.3235
diM-Ether (lbmole/hr)	0.2421	0.0000
Methanol (lbmole/hr)	476.6885	2.9112

Flow Basis: Molar The composition option is selected

Product Recoveries

	Methanol Out	Wastewater Out
Flow Rate (lbmole/hr)	507.7195	582.2347
	---	---
H2O (%)	5.0464	94.9536
diM-Ether (%)	100.0000	0.0000
Methanol (%)	99.3930	0.6070

COLUMN PROFILES

Reflux Ratio: 1.019 Reboil Ratio: 1.455 The Flows Option is Selected Flow Basis: Molar

Column Profiles Flows

	Temp (F)	Pres (psia)	Net Liq (lbmole/hr)	Net Vap (lbmole/hr)	Net Feed (lbmole/hr)	Net Draws (lbmole/hr)	Duty (Btu/hr)
Condenser	269.2	124.7	517.2	---	---	507.7	1.35e+007
1__Main Tower	270.7	124.7	511.7	1025	---	---	---
2__Main Tower	272.0	124.9	506.3	1019	---	---	---
3__Main Tower	273.3	125.0	501.0	1014	---	---	---
4__Main Tower	274.7	125.1	495.9	1009	---	---	---
5__Main Tower	276.0	125.2	490.9	1004	---	---	---
6__Main Tower	277.4	125.3	486.2	998.7	---	---	---
7__Main Tower	278.8	125.4	481.6	993.9	---	---	---
8__Main Tower	280.2	125.6	477.2	989.3	---	---	---
9__Main Tower	281.5	125.7	473.1	984.9	---	---	---
10__Main Tower	282.9	125.8	469.3	980.8	---	---	---
11__Main Tower	284.1	125.9	465.9	977.1	---	---	---
12__Main Tower	285.3	126.0	462.9	973.7	---	---	---
13__Main Tower	286.4	126.2	460.4	970.7	---	---	---
14__Main Tower	287.4	126.3	458.2	968.1	---	---	---
15__Main Tower	288.2	126.4	456.3	965.9	---	---	---
16__Main Tower	288.9	126.5	454.9	964.1	---	---	---
17__Main Tower	289.5	126.6	453.7	962.6	---	---	---
18__Main Tower	290.1	126.7	452.1	961.4	---	---	---
19__Main Tower	291.3	126.9	1515	959.8	1090	---	---
20__Main Tower	290.7	127.0	1513	932.8	---	---	---
21__Main Tower	290.8	127.1	1512	930.7	---	---	---
22__Main Tower	291.0	127.2	1512	930.2	---	---	---
23__Main Tower	291.4	127.3	1510	929.4	---	---	---
24__Main Tower	291.9	127.5	1508	928.0	---	---	---
25__Main Tower	292.9	127.6	1503	925.4	---	---	---
26__Main Tower	294.5	127.7	1496	920.9	---	---	---
27__Main Tower	297.4	127.8	1482	913.3	---	---	---
28__Main Tower	304.8	127.9	1457	899.7	---	---	---
29__Main Tower	314.5	128.0	1441	874.4	---	---	---
30__Main Tower	325.5	128.2	1431	858.5	---	---	---
31__Main Tower	335.2	128.3	1429	848.9	---	---	---
32__Main Tower	341.3	128.4	1429	846.7	---	---	---
Reboiler	344.8	128.4	---	847.0	---	582.2	1.34e+007

Column Profiles Energy

	Temperature (F)	Liq Enthalpy (Btu/lbmole)	Vap Enthalpy (Btu/lbmole)	Heat Loss (Btu/hr)
Condenser	269.2	-9.858e+004	-8.489e+004	---
1__Main Tower	270.7	-9.937e+004	-8.538e+004	---
2__Main Tower	272.0	-1.002e+005	-8.571e+004	---
3__Main Tower	273.3	-1.009e+005	-8.603e+004	---
4__Main Tower	274.7	-1.017e+005	-8.634e+004	---
5__Main Tower	276.0	-1.024e+005	-8.664e+004	---
6__Main Tower	277.4	-1.032e+005	-8.694e+004	---
7__Main Tower	278.8	-1.039e+005	-8.723e+004	---
8__Main Tower	280.2	-1.046e+005	-8.751e+004	---

9__Main Tower	281.5	-1.053e+005	-8.778e+004	---
10__Main Tower	282.9	-1.059e+005	-8.803e+004	---
11__Main Tower	284.1	-1.065e+005	-8.827e+004	---
12__Main Tower	285.3	-1.070e+005	-8.848e+004	---
13__Main Tower	286.4	-1.075e+005	-8.867e+004	---
14__Main Tower	287.4	-1.079e+005	-8.884e+004	---
15__Main Tower	288.2	-1.082e+005	-8.898e+004	---
16__Main Tower	288.9	-1.085e+005	-8.910e+004	---
17__Main Tower	289.5	-1.087e+005	-8.920e+004	---
18__Main Tower	290.1	-1.088e+005	-8.928e+004	---
19__Main Tower	291.3	-1.090e+005	-8.933e+004	---
20__Main Tower	290.7	-1.090e+005	-8.935e+004	---
21__Main Tower	290.8	-1.091e+005	-8.937e+004	---
22__Main Tower	291.0	-1.091e+005	-8.942e+004	---
23__Main Tower	291.4	-1.092e+005	-8.949e+004	---
24__Main Tower	291.9	-1.094e+005	-8.961e+004	---
25__Main Tower	292.9	-1.097e+005	-8.984e+004	---
26__Main Tower	294.5	-1.102e+005	-9.022e+004	---
27__Main Tower	297.4	-1.110e+005	-9.088e+004	---
28__Main Tower	304.8	-1.130e+005	-9.194e+004	---
29__Main Tower	314.5	-1.146e+005	-9.471e+004	---
30__Main Tower	325.5	-1.160e+005	-9.710e+004	---
31__Main Tower	335.2	-1.169e+005	-9.925e+004	---
32__Main Tower	341.3	-1.173e+005	-1.006e+005	---
Reboiler	344.8	-1.175e+005	-1.014e+005	---

FEEDS / PRODUCTS

Flow Basis: Molar

	Stream	Type	Duty	State	Flows	Enthalpy	Temp
			(Btu/hr)		(lbmole/hr)	(Btu/lbmole)	(F)
Condenser	Q Condense 2	Energy	1.4e+007		---	---	---
	Methanol Out	Draw	---	Liquid	508	-9.9e+004	269.17

1__Main Tower

2__Main Tower

3__Main Tower

4__Main Tower

5__Main Tower

6__Main Tower

7__Main Tower

8__Main Tower

9__Main Tower

10__Main Tower

11__Main Tower

12__Main Tower

13__Main Tower

14__Main Tower

15__Main Tower

16__Main Tower

17__Main Tower							
18__Main Tower							
19__Main Tower	Methanol and Water Outlet	Feed	---	Liquid	1.09e+003	-1.1e+005	305.21
20__Main Tower							
21__Main Tower							
22__Main Tower							
23__Main Tower							
24__Main Tower							
25__Main Tower							
26__Main Tower							
27__Main Tower							
28__Main Tower							
29__Main Tower							
30__Main Tower							
31__Main Tower							
32__Main Tower							
Reboiler	Q Reboil 2	Energy	1.3e+007	---	---	---	
	Wastewater Out	Draw	---	Liquid	582	-1.2e+005	344.85

PERFORMANCE SUMMARY FOR INTERNAL OPTION: Internals-1@Main Tower@COL2

Number Of Stages	32
Total Height (ft)	64.00
Total Head Loss (in)	133.8
Total Pressure Drop (inH2O(60F))	99.06
Number Of Sections	1
Number Of Diameters	1
Pressure Drop Across Sump (psi)	---

Section	Start	End	Height (ft)	Diameter (ft)	Internals Type	Tray or Packing Type	Section Pressure Drop (inH2O(60F))	Approach To Flood (%)	Limiting Stage
CS-1	1__Main Tower	32__Main Tower	2.000	3.000	Trayed	Sieve	99.06	79.66	

SETUP

Sub-Flowsheet

Internal Feed Stream	External Feed Stream	Transfer Basis
Q Reboil 2	Q Reboil 2	None Req'd
Methanol and Water Outlet	Methanol and Water Outlet	P-H Flash

Internal Prod Stream	External Prod Stream	Transfer Basis
Q Condense 2	Q Condense 2	None Req'd
Methanol Out	Methanol Out	P-H Flash
Wastewater Out	Wastewater Out	P-H Flash

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
Q Reboil 2		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
Q Condense 2		
Methanol Out		

DYNAMICS

Vessel Dynamic Specifications

Vessel	Reboiler @COL2	Condenser @COL2
Diameter (ft)	3.914	3.914
Height.0 (ft)	5.871	5.871
Volume.0 (ft3)	70.63	70.63
Liquid Volume Percent (%)	50.00	50.00
Level Calculator	Horizontal cylinder	Horizontal cylinder
Fraction Calculator	Use levels and nozzles	Use levels and nozzles
Vessel Delta P (psi)	0.0000	0.0000
Fixed Vessel P Spec (psia)	128.4	124.7
Fixed P Spec Active	Not Active	Not Active

Other Equipment in Column Flowsheet

Holdup Details

	Pressure (psia)	Volume (ft3)	Bulk Liquid Volume (ft3)
Condenser	124.7	38.60	36.36
1__Main Tower	124.7	34.32	0.4097
2__Main Tower	0.0000	0.0000	0.0000
3__Main Tower	0.0000	0.0000	0.0000
4__Main Tower	0.0000	0.0000	0.0000
5__Main Tower	0.0000	0.0000	0.0000
6__Main Tower	0.0000	0.0000	0.0000
7__Main Tower	0.0000	0.0000	0.0000
8__Main Tower	0.0000	0.0000	0.0000
9__Main Tower	0.0000	0.0000	0.0000
10__Main Tower	0.0000	0.0000	0.0000
11__Main Tower	0.0000	0.0000	0.0000
12__Main Tower	0.0000	0.0000	0.0000
13__Main Tower	0.0000	0.0000	0.0000
14__Main Tower	0.0000	0.0000	0.0000
15__Main Tower	0.0000	0.0000	0.0000
16__Main Tower	0.0000	0.0000	0.0000
17__Main Tower	125.4	34.32	0.6126

18__Main Tower	0.0000	0.0000	0.0000
19__Main Tower	0.0000	0.0000	0.0000
20__Main Tower	0.0000	0.0000	0.0000
21__Main Tower	0.0000	0.0000	0.0000
22__Main Tower	0.0000	0.0000	0.0000
23__Main Tower	0.0000	0.0000	0.0000
24__Main Tower	0.0000	0.0000	0.0000
25__Main Tower	0.0000	0.0000	0.0000
26__Main Tower	0.0000	0.0000	0.0000
27__Main Tower	0.0000	0.0000	0.0000
28__Main Tower	0.0000	0.0000	0.0000
29__Main Tower	0.0000	0.0000	0.0000
30__Main Tower	0.0000	0.0000	0.0000
31__Main Tower	0.0000	0.0000	0.0000
32__Main Tower	126.2	34.32	0.4627
Reboiler	126.2	4948	38.88

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